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NEW DEVELOPMENTS IN AEROBIC TREATMENT OF WASTEWATER

May 21, 1986

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NEW DEVELOPMENTS IN AEROBIC TREATMENT OF WASTEWATER

Proceedings of a Seminar held at Hart House,
University of Toronto

MAY 21, 1986

Sponsored and Organized by:

POLLUTION CONTROL ASSOCIATION OF ONTARIO

and the

ONTARIO MINISTRY OF THE ENVIRONMENT

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NEW DEVELOPMENTS IN ACTIVATED SLUDGE PROCESS

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1. INTRODUCTION

Although the activated sludge process has been the basic biological wastewater treatment process for about seven decades, it is still a preferable treatment method for many applications, competing easily with other processes due to its special features (Ganczarczyk, 1983). In the recent years the scientific basis for the process has been much better understood and many substantial advances in the process technology and equipment occured. It is impossible to present in this paper all the new developments of the process. Therefore, the discussion will be limited to those developments which appear to be most important and which are expected to have a lasting influence on the process application.

In a general way it is possible to divide the new developments in the activated sludge treatment into those which involve directly the process, and into those which lead to some process modifications through changes in the reactor design. In the scope of the first group, the following items have been selected for a closer examination: bio-augmentation, application of biomass carriers, and biomass retention in aeration tanks. As the second group the non-continuous systems (sequencing batch reactors and continuously fed but intermittently decanted units), intrachannel clarification, compartmentization of aeration tanks, and German tower reactors, will be discussed (Table I).

TABLE I. SELECTED PROCESS AND REACTOR DEVELOPMENTS

A. Direct Process Developments:

- bioaugmentation;

- application of biomass carriers:
  - micro-carriers (e.g. PACT process),
  - macro-carriers (e.g. CAPTOR/LINPOR systems).

- biomass retention:
  - packed media in aeration tanks,
  - CAPTOR/LINPOR systems,
  - SURFACT system

B. Reactor-Process Interactions:

- non-continuous systems:
  - sequencing batch reactors (SRBs),
  - continuously fed and intermittently decanted systems (CFIDs).

- interchannel clarification,

- compartmentization of aeration tanks,

- German tower reactors

The broad field of biological nutrients removal will be not dealt with here, as it should be presented in a form of a separate paper. Similarly, the important developments in the evaluation of aeration equipment are discussed elsewhere (Ganczarczyk, 1985).

2. BIOAUGMENTATION

Bioaugmentation in the activated sludge process is based on additions of some special microorganisms to the biomass generated in the specific wastewater treatment system by normal procedures. The purpose of such additions is to improve the treatment, speed-up the process starting procedures, make easier to get out from some process upset conditions, and to reduce sludge production or improve sludge handling. As bioaugmentation products, usually a liquid culture or a freeze-dry biomass is being used.

In some reports the effectiveness of bioaugmentation has been emphasized, but some critical voices on the subject have also been heard. The survival and a dynamic growth of the bioaugmentation bacteria may be greatly influenced by the antagonistic action of other competing microbes. Presently marketed products, both dry and liquid cultures, are recommended for treatment of various industrial wastewater from those containing cyanide, phenols, sulfur compound to those with high concentreation of grease, protein and starches. About 60 different bioaugmentation products are available throughout the world.

It may be expected that recent progress of bioengineering in medicine and agriculture, will make possible an application of

microorganisms with recombinant DNA for treatment of some industrial effluents and perhaps even sewage. The bacteria with recombinant DNA may be more resistant to higher and fluctuating loadings and show better flocculating abilities. In several laboratories there are studies carried out on the subject. However, it is difficult presently to judge whether these technologies will prove enough practical and safe.

3. BIOMASS CARRIERS

There are many unanswered questions associated with activated sludge treatment of wastewater in the presence of solid particles which may affect the process in many different ways. Surface of these particles may be a supporting medium for some selected microorganisms and it is possible that the this type of growth may be more physiologically active. The carrier particles may also possess different sorption characteristics which may differently affect the process kinetics by increasing local concentration of the metabolic substrates and nutrients, and by storing the excess of the substrates in cases of shock loadings. It was experimentally shown (Lu and Ganczarczyk, 1983) that application of biomas carriers allows for maintaining a higher biomass concentration in aeration tanks, helps to control sludge volume index and improves substrate removal rates.

The most common biomass carrier in the activated sludge process is the powdered activated carbon process. It may be claimed that the process modification based on powdered activated carbon addition to the aeration tanks, and called the PACT system, is the most important improvement in the activated sludge treatment

which occured in the last quarter of the century.

4. RETENTION OF BIOMASS IN AERATION TANKS

The retention of some amounts of biomass in aeration tanks may be beneficial for various treatment systems prone to the sludge bulking phenomena or operating at high mixed liquor suspended solids level. In the latter case the secondary clarifiers would be less subjected to the excessive sludge loadings.

The idea of biomass retention in aeration tanks has originated as the old concept of the submerged contact aeration system, but is recently reborn as a design of the aeration tanks containing plastic media packings below aeration devices, and as an addition of plastic foam particles to some aeration tanks. The latter idea has developed separately in the U.K. and Germany. The British approach, called CAPTOR process, was invented by B. Atkinson. A possibility was shown that the biomass would develop only on the carrier making the secondary clarifier superfluous. The biomass could be wasted by squeezing it from the support particles. In the German technology, called LINPOR process, the plastic foam particles are added only as supplementary means to improve performance of overloaded plants (Hegemann, 1984).

The SURFACT process (Guarino, 1983) can also be considered as a method for biomass retention in aeration tanks. This method involves a retrofitting of the conventional, compressed air-aerated, aeration tanks with a special type of rotating biological discs. These discs are equipped with peripheral cups which capture the diffused air-bubbles and thus provide energy

for the discs rotation. The biofilm on the discs increases the biomass inventory of the system and the aeration capacity of the discs contributes to the total aeration capacity of the treatment unit.

5. SEQUENCING BATCH REACTORS

Application of the sequencing batch reactors (SBR) method for the activated sludge process represents a break with the continuous flow treatment concept of Jones (1914) and a return to the idea of an early development of the process: fill-and-draw reactors. In the recent applications, the sequencing batch reactors are used mostly for extended aeration systems, but this approach is not limited to low F/M ratios and high values of sludge retention times. It is expected that the introduction of the SBRs may result in up to 20 percent savings in some plant construction costs and in up to 10 percent savings in the power requirements.

A SBR system may include one or more tanks, each having five basic modes of operation: fill, react, settle, draw, idle. Fill is the receiving of the wastewater; react describes the period of treatment; settle is clarification; draw is effluent discharge; and idle is the waiting period until the next cycle begins. Time clock controls can be used, as well as level sensors or turbidity sensors, to operate the system. Microprocessors are also easily adapted to control SBR operations. Solids wasting is done after the settle period or during react and can be accomplished from twice per month for a simple one-tank system to once-per-cycle for a more complex

multi-tank system. Neither primarily nor separate secondary clarifiers are required, and no recycle pump is necessary, indicating potential for significant savings in construction as well as in operation and maintenance costs.

More important recent references to this technology cover papers by Arora (1985), Irvine (1985), Irvine et al. (1985), Manning and Irvine (1985), etc.

An interesting modification of the above approach are applications of continuously fed but intermittently decanted systems (CFID),developed in Australia and described and discussed by Goronszy (1979). The latter are simpler than SBRs but may be not able to produce equally good effluent due to leakage of incoming wastewater during settling and decanting portions of the treatment cycle.

6. INTRACHANNEL CLARIFICATION

By locating the final clarifier in the oxidation ditch type aeration tanks (internal recycle), the activated sludge process may be substantially simplified through elimination of a separate external recycle system. This approach was originally tested by Pasveer in 1957 with a limited success, but recently was developed into a viable design by the Burns and McDonnel Engineers (Christopher et al., 1984). The latter design, called BMTS, uses an innovative clarifier which consists of front and rear baffle walls across the aeration channel, forcing the horizontal flow underneath a quiescent zone within the clarifier. The bottom panels of the clarifier distribute the up-flow movement of the clarified effluent, and solids are returned back

to the horizontal flow. For aeration in the BMTS system originally a down-pumping, submerged-turbine aerator with a U-tube discharge was used, but later a diffused aeration with a forced horizontal mechanical mixing proved to be more advantageous.

7. COMPARTMENTIZATION OF AERATION TANKS

Compartmentization of aeration tanks (German misnomer: cascading) simulates plug-flow conditions in aeration tanks. It is of a special importance for some process modifications effective in the biological nutrients removal. These topics are outside the scope of this paper.

8. TOWER REACTORS

The scarcity of space and the necessity to avoid air pollution associated with some chemicals stripping from aeration tanks led to the development of the tower activated sludge reactors for treatment of industrial effluents in Germany. This development combines high oxygen transfer rates with minimizing the air use, and is connected with application of aeration devices which optimum depth of operation is quite substantial.

The German tower reactors were developed as two somewhat different designs: the Bayer tower and the Bio-Hoch reactor by Hoechst Co. The Bayer tower utilizes special Bayer aeration nozzles causing a spiral movement of the mixed liquor. At the Leverkusen plant, the Bayer tower operates with the water depth of 26 m and 75-80% of the oxygen from the air is absorbed. The Bio-Hoch reactor in the Merck plant in Darmstadt operates with

the water depth of 18 m and with draft tubes over the aeration nozzles. The reported oxygen utilization is in the range between 60 and 70%. It is not likely that these designs might become popular elsewhere except of locations with a similar concentration of industry.

9. CONCLUSION

It may be expected that the future advances in the activated sludge process will follow the three basic directions:

- further development of technological steps to increase the effectiveness and stability of the treatment.
- development of control sensors and control systems to guarantee the possibly reliable treatment; and
- development of more efficient aeration systems to decrease the energy requirements of the process and thus make it more economical.

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OPERATION OF AN ULTRA-HIGH RATE ACTIVATED SLUDGE SYSTEM TREATING TANNERY WASTE

By J.W.G. Rupke[1], H.J. Knibbe[2]

INTRODUCTION

Production facility

Collis Leather, a division of Canada Packers Limited operates a leather tannery in the Town of Aurora, Ontario. The tannery opened its doors for operation in 1911 and was acquired by Canada Packers Limited in 1936.

The tannery manufactures side and calf leathers processing up to 6,000 sides per week. Processing is carried out from the brine cured hide, as received from the slaughter houses to a finished leather ready for manufacture into leather goods. The manufacturing of chrome tan leather can be separated into two phases: (i) the "Wet End Processes" which are completed in solution in large rotating drums, and (ii) the "Dry End Processes" such as drying, buffing, finishing and grading. The Wet End Processes generates the great majority of the organic and hydraulic loadings to the wastewater treatment plant.

West End Processes

1) Hide Washing and Fleshing

The brine cured hide is first washed to remove dirt and manure. Following this, the hide is mechanically fleshed to remove excess fat and muscle. The hide is then allowed to soak to allow the replacement of natural moisture which was lost during curing.

2) Beaming

The hide is now ready to have the hair removed. Beaming or dehairing is done chemically with the addition of lime, sodium sulfhydrate and other agents all in solution. These chemicals free and dissolve the hair material along with some soluble skin proteins.

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MOE/PCAO Seminar (May 21, 1986)

3) Tanning

The tanning process is composed of 3 steps: (i) Bating, (ii) Pickling and (iii) Chrome Tanning. Bating utilizes select enzymes which act on the hide proteins resulting in improved leather texture and softness Ammonium sulphate is added in this stage to remove any remaining sulfides, lime and insoluble calcium soaps. In the pickling operation the pH of the hide is reduced in preparation for receiving the tanning solutions. Sulphuric acid and other agents are added to drive the solution to a pH of 2-3.

The chrome tanning step is used to make the leather resistant to bacterial decay and heat shrinkage. In this stage trivalent chromium is added, which penetrates the hide and forms a stable complex with the proteinaceous material.

In the normal chrome tan process approximately 60-70% of the chromium is incorporated into the hide. Collis Leather utilizes a high exhaustion process called Baychrome which allows better than 95% of the trivalent chromium to be incorporated into the hides. This significantly reduces the amount of chromium which will require removal by treatment.

4) Retan, Colouring and Fatliquoring

To improve hide quality or provide specific characteristics the hides may be retanned using synthetic or vegetable extract tanning compounds after completion of chromium tanning.

The colouring is achieved by the addition of synthetic dyes in an acidic solution.

In fatliquoring various oils are added to improve the pliability of the leather and improve its tensile strength.

Dry End Processes

The Dry End Processes consist of drying, conditioning, buffing, finishing and sorting of the hides. These procedures contribute very little to the organic waste loadings. Cooling and vacuum water in this area can contribute significantly to the hydraulic loading, however, Collis Leather has instituted water recycle reducing the total water usage by an estimated 40-45%.

Treatment Facility

All production wastewater is collected in the main process wastewater sump. This wastewater is pumped directly from the main sump to the treatment facility. The production plant has two other major sumps utilized to store various wastewater streams. Wastewater discharges from these sumps are staggered throughout the day to provide some equalization of the organic loadings.

Sulfide oxidation is provided within the production plant on all major sulfide bearing wastewater streams prior to discharge to the treatment facility.

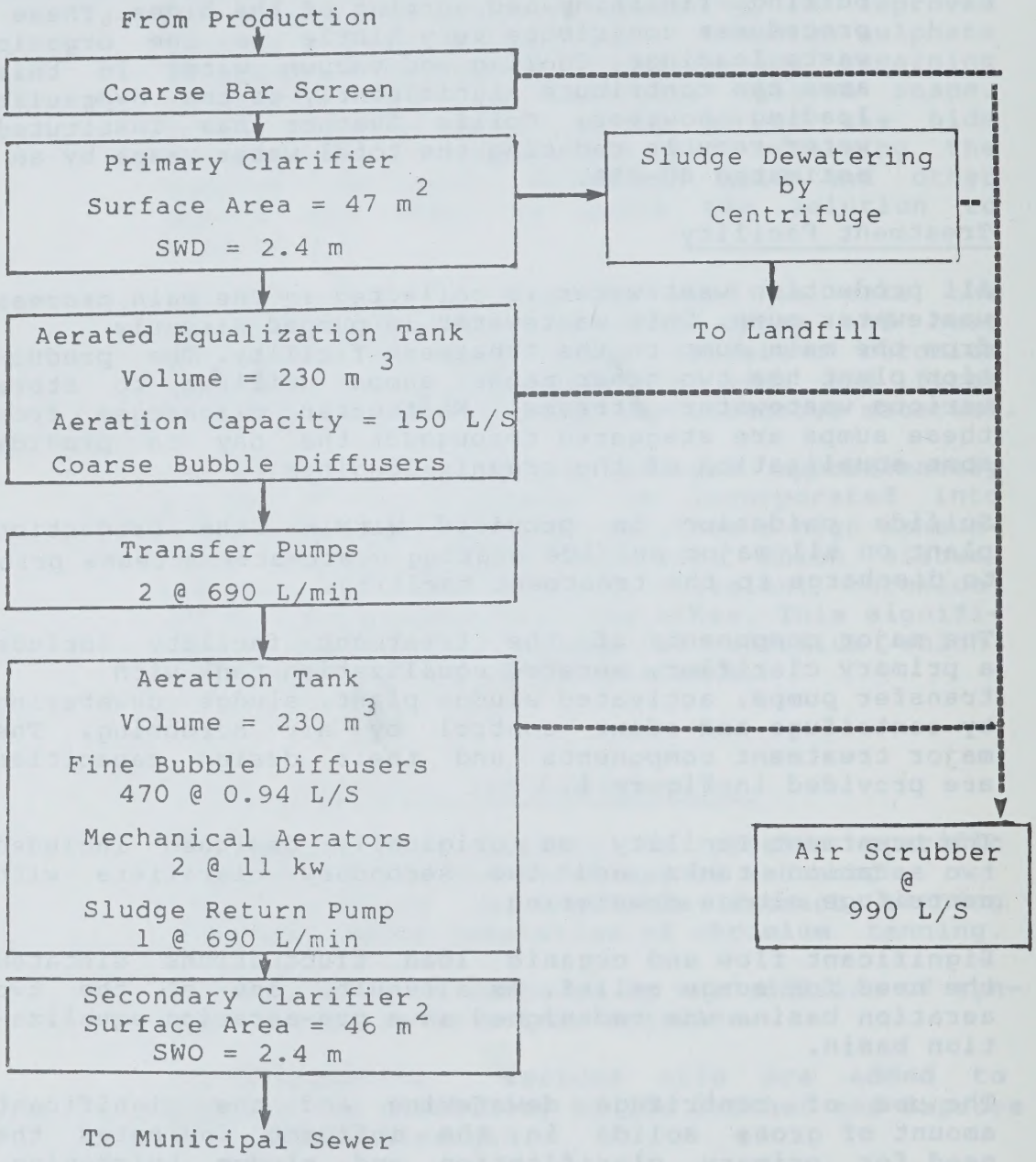
The major components of the treatment facility include a primary clarifier, aerated equalization tank with transfer pumps, activated sludge plant, sludge dewatering by centrifuge and odour control by air scrubbing. The major treatment components and their design capacities are provided in Figure 1.

The treatment facility as originally designed included two aeration tanks and two secondary clarifiers with centrifuge sludge dewatering.

Significant flow and organic load fluctuations dictated the need for surge relief. As a result, one of the two aeration basins was redesigned as a pre-aeration equalization basin.

The use of centrifuge dewatering and the significant amount of gross solids in the influent indicated the need for primary clarification and sludge thickening. Since a large excess in secondary clarifier capacity was available one unit was converted into a primary clarifier. The primary clarifier provides for the removal of gross solids and allows blending of primary and secondary

FIGURE 1 - SCHEMATIC OF WASTEWATER TREATMENT FACILITY



Treatment Facility cont'd

sludges as well as sludge thickening.

Coarse screening is provided to remove gross solids and large scraps of leather which might interfere with pumping and sludge dewatering. After screening the wastewater is treated by primary clarification and is then discharged to the equalization tank. It would be preferable to have equalization prior to primary clarification, however, hydraulic restriction would not allow this flow pattern without additional pumping.

Wastewater is pumped from the equalization tank to the aeration tank. The pump flow rate is set manually by the operator using adjustable cycle timers. Floats provide automatic pump shut-off when the equalization basin is empty. Coarse bubble aeration is supplied to provide mixing and pre-aeration.

The aeration basin has a volume of 230 m^3 and a side water depth of 2.4 meters. Aeration is provided using 470 fine bubble diffuser domes each discharging 0.94 L/second and two 11 Kw low speed surface aerators

The secondary clarifier has a surface area of 46 m^2 and a side water depth of 2.4 meters. The activated sludge return pump has a capacity of 680 L/min or 200% of average daily flow.

Phosphorous as a nutrient is added to the aeration basin batchwise on a daily basis.

Sludge dewatering is provided by one Sharples P-3000 solid bowl sludge dewatering centrifuge which has a nominal capacity of 70 L/min and a maximum bowl speed of 3250 rpm. The centrifuge has been retrofitted with a Sharples "low polymer use scroll" to reduce polymer consumption.

The production facility is located in the heart of the Town of Aurora, the nearest residence being located within 45 metres of the treatment facility itself. In order to minimize odour emissions an air scrubbing system was incorporated into the treatment plant design. All the air discharged from the surge and aeration tanks as well as air drawn from the centrifuge and screening rooms is passed through the air scrubber prior to discharge. Sodium hypochlorite is added to the scrubbing solution

Treatment Facility cont'd

to strip out odour causing compounds. The sodium hypochlorite concentration in the recycle solution is controlled automatically using a continuously recording ORP monitoring system.

OPERATING RESULTS

The sewage characteristics have not varied significantly over the five years of operating data accumulated at this facility. There are, however, seasonal variations in sewage characteristics related primarily to the cleanliness of the raw hides being received at the tannery. Local winter hides tend to carry considerably more dirt and fecal matter than the hides from warmer climates.

The quantity of sewage produced relates directly to the level of production in the Tannery.

Table 2-1 shows the normal range of raw sewage characteristics and flow rates.

The primary clarifier removes 32% of the BOD and 25% of the suspended solids.In addition the primary allows for the concentration of waste activated sludge and raw sludge storage prior to dewatering by centrifuge. Primary effluent quality is shown in Table 2-2. It should be noted that on average the primary effluent is deficient in phosphorous based on a 100:1 ratio of BOD:phosphorous. The phosphorous level is suppl mented to ensure adequate nutrients are present for the operation of the activated sludge system. Prior to initiating the phosphorous addition, periodic episodes of poor sludge settling were experienced related to lack of available phosphorous. Table 2-3 shows the activated sludge operating conditions. The very high volumetric loading rates 5.5 Kg/m^3 /day (320 lb BOD/1000 ft^3/day) and the high MLSS levels all contribute to the extremely high sludge respiration rate (250 mg/L/hr). This high rate of oxygen demand is just barely met by the two mechanical surface aerators and the fine bubble diffused air system. The secondary clarifier has an hydraulic surface loading rate of 10.8 m^3/m^2 /day (220 gal/day/ft^2) at average flow. Actual peak hydraulic loadings are 21.6 m^3/m^2/day (440 gal/day/ft^2) The secondary effluent quality is shown in Table 2-4. The sewer use bylaw limits for this effluent are BOD-500 mg/L; SS 600 mg/L and chromium 5 mg/L.

The effluent quality meets the sewer use bylaw at all times for SS and total chromium The effluent BOD meets

TABLE 2-1

COLLIS LEATHER

RAW SEWAGE CHARACTERISTICS

CHARACTERISTIC	MINIMUM	MAXIMUM	AVERAGE	STANDARD DEVIATION
BOD mg/L	1010	5820	3415	1480
COD mg/L	4000	14000	8400	2875
Suspended Solids mg/L	420	3370	1490	820
NH_3 mg/L	100	1125	405	295
PO_4 as P mg/L	2.7	47	16.7	11.3
Total Chromium mg/L	2.8	25.0	10.7	5.2
pH	3.84	11.6	9.14 (median)	-----
Flow Rate m^3/day	230	90	500	-----

TABLE 2-2

COLLIS LEATHER

PRIMARY EFFLUENT CHARACTERISTICS

CHARACTERISTIC	MINIMUM	MAXIMUM	AVERAGE	STANDARD DEVIATION
BOD mg/L	675	4180	2310	875
COD mg/L	2250	9700	5840	1755
Suspended Solids mg/L	250	7040	1120	1374
NH_3 mg/L	270	2280	675	485
NO_3 as N mg/L	20	900	193	220
PO_4 as P mg/L	3.3	47	17.3	12
Total Chromium mg/L	1.5	33	8.4	7.4
pH	7.34	9.65	8.89 (median)	----

TABLE 2-3

ACTIVATED SLUDGE OPERATING CONDITIONS

	OPERATING RANGE	AVERAGE
MLSS mg/L	7000-12000	10000
F/M	0.3 - 0.75	0.55
30 min settling (%sludge)	50 - 96	75
SVI	70 - 140	90
Organic Loading $kg/m^3/day$ (lb BOD/1000 ft^3/day	3.2 - 6.4 200 - 400	5.1 320
Recycle rate % of Q	150 - 250	200
Flow rate Q m^3/day	227 - 910	500
DO mg/L	0.2 - 1.5	0.6
Respiration Rate mg/L/hr	100 - 400	250
SRT days	2.0 - 4.0	3.2
Activated Sludge Production Kg/Kg BOD_{rem}	---- ---	0.67

TABLE 2-4

COLLIS LEATHER

SECONDARY EFFLUENT CHARACTERISTICS

CHARACTERISTIC	MINIMUM	MAXIMUM	AVERAGE	STANDARD DEVIATION
BOD mg/L	90	1300	420	335
COD mg/L	550	4400	1770	870
Suspended Solids mg/L	20	435	145	97
NH_3 mg/L	250	1300	650	306
NO_3 as N mg/L	4	50	21	12
PO_4 as P mg/L	0.6	13.3	5.2	3.7
Total Chromium mg/L	0.2	1.5	0.8	0.4
	7.63	8.21	7.95 (median)	-----

OPERATING RESULTS cont'd

the bylaw on average, however, during periods of heavy organic loading the treatment capacity can be exceeded and the effluent BOD will then rise above the bylaw limit. Figure 2-1 shows the relationship between the volumetric organic loading rate and the effluent quality. The fitted curve shows that if the volumetric loading exceeds 6.6 $Kg/m^3/day$ (412 $lb/1000ft^3/day$) the effluent quality will exceed 500 mg/L BOD. This limitation is likely due to the limited oxygen supply available and should not be considered a limitation of the process.

Close examination of the effluent data shows that the nitrate levels decrease from 193 mg/L to 21 mg/L as the wastewater is treated. The denitrification occurs due to the limited oxygen supply available and occasionally results in floating sludge in the final clarifier due to nitrogen gas generation in the stored sludge in the final clarifier.

The chromium is removed via the sludge stream as a precipitated chromium hydroxide. All the chromium is in the reduced trivalent state with no detectable amounts of the toxic hexavalent chromium. The activated sludge system shows no evidence of chromium toxicity even with 1600 mg Cr^{+3}/Kg of dry solids in the mixed liquor.

OXYGEN SUPPLY

The fine bubble diffuser system requires monthly cleaning using HCl gas to remove the biological and inorganic fouling on the surface of the ceramic domes. This system of gas cleaning has proven satisfactory to maintain the system backpressure at the original values and the oxygen transfer at acceptable levels.

Steady state oxygen transfer tests have shown that the combine oxygen supply is capable of supplying up to 60 Kg O_2/hr. Measurements of the Beta factor show that because of the high dissolved solids the Beta values are in the 0.75 to 0.78 range. High operating temperatures (30^0C) further limit the capacity of the oxygen supply system.

It is estimated that the mechanical aerators can supply 0.85 $KgO_2/kw/hr$ giving a total supply of 19 KgO_2 /hr from the mechanical aerators. The remaining 41 KgO_2 /hr comes from the fine bubble diffuser system. The

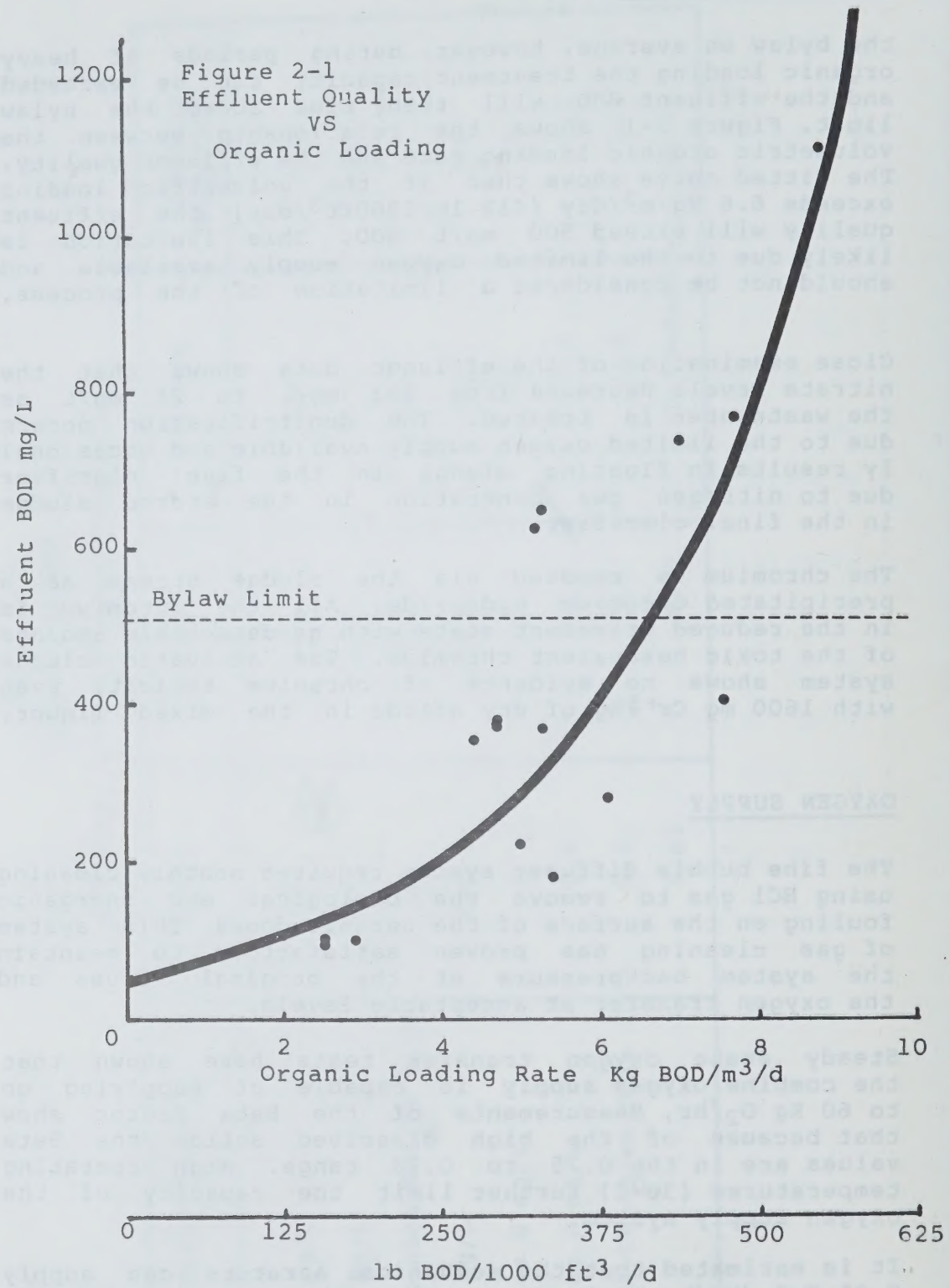


Figure 2-1
Effluent Quality
VS
Organic Loading
Effluent BOD mg/L
1200
1000
800
600
400
200
0
Bylaw Limit
2
4
6
8
10
Organic Loading Rate Kg BOD/m3/d
0
125
250
375
500
625
lb BOD/1000 ft3 /d

OXYGEN SUPPLY cont'd

oxygen transfer efficiency is estimated to be 9.3% based on an oxygen input of 442 KgO_2/hr. Considering the adverse conditions that the oxygen supply system is exposed to (high temperatures, low Alpha and Beta factors and only 7 ft submergence) the fine bubble aeration system is performing well.

SLUDGE DEWATERING

The waste treatment facility generates an average of 1200 Kg/day of dry solids, of which approximately half is primary sludge and the remainder is waste activated sludge. The blended raw sludge fed to the centrifuge averages 7.0% total solids. On a blended sludge the centrifuge will generate a sludge cake averaging 25% total solids, however, this will drop as low as 17% when activated sludge alone is dewatered.

The centrifuge centrate averages 1875 mg/L of suspended solids providing a percent recovery rate of 97% across the centrifuge.

Polymer usage for dewatering averages 5.2 Kg/tonne of dry solids.

Although the cake shows significant levels of trivalent chromium leachability testing has shown it to be acceptable for disposal at a municipal landfill site. The analytical characteristics of raw and dewatered sludge are summarized in Table 2.5.

CONCLUSIONS

The activated sludge system can be operated, far above what is considered the normal operating range of organic loadings. The effluent quality is suitable for discharge to a municipal sewer system. BOD removal efficiency in the 80% to 85% range can be achieved at organic volumetric loading rates ten time those considered normal. The system tends to fail due to inadequate oxygen supply or failure of the solids separation system. Sludge production is not abnormally high and dewaters in a centrifuge to produce a cake acceptable at a landfill site. Both the slump test and the leachability requirements can be consistently met.

TABLE 2-5

COLLIS LEATHER
SLUDGE CHARACTERISTIC

		MINIMUM	MAXIMUM	AVERAGE	STANDARD DEVIATION
Raw Sludge	Total Solids %	5.4	12.4	7.0	1.7
	Total Chromium mg/Kg of wet cake	160.0	760.0	374.0	194.0
Sludge Cake	Total Solids %	20.6	32.4	24.6	2.8
	Total Chromium mg/Kg of wet cake	890.0	5000.0	1940.0	1100.0
Centrate	Suspended Solids	425.0	7500.0	1875.0	1715.0

CONCLUSIONS cont'd

Chromium removal is almost complete with all the chromium being incorporated into the sludge stream.

Sutton WPCP: A Low Cost Nitrification-Denitrification System for the Control of Ammonia and Hydrogen Sulphide in Waste Stabilization Pond Effluents

Introduction and Background

Facultative waste stabilization ponds in Ontario are usually void of dissolved oxygen during periods of ice and snow cover as these conditions prevent the replenishment of oxygen from algal photosynthesis and surface aeration. Due to the low oxygen and nitrate concentration in ponds at this time of year, the oxygen requirements of heterotrophic bacteria are satisfied by sulphates. The biochemical reduction of sulphate results in the formation of hydrogen sulphide and other sulphide compounds. In addition, an increase in ammonia-nitrogen may occur due to the conversion of organic nitrogen to ammonia under anaerobic conditions.

As a result, the ammonia and hydrogen sulphide concentrations in pond effluents may exceed levels acceptable for discharge to receiving waters. This is of particular concern with continuously discharging ponds. It could, however, also be a problem in meeting effluent criteria during spring discharge of seasonal retention ponds. Annual retention ponds, if discharged during the fall, usually produce effluents of acceptable quality. All three types of waste stabilization ponds, however, may cause odour problems during the spring ice breakup period.

To control both ammonia and hydrogen sulphide in pond effluents, particularly where annual retention cannot be provided, the former Wastewater Treatment Section of the Pollution Control Branch of the Ministry of the Environment investigated the design and operation of a pre-treatment unit which produces a highly nitrified effluent that can be discharged into an adjacent pond. The pre-treatment unit itself consists of a complete-mix aeration cell followed by clarification and sludge return. Pilot studies were carried out at the Ontario Experimental Facility at Brampton during the winter of 1980/81. By testing this treatment system under varied operational conditions, it was discovered that increased sludge age and controlled release of sludge from the clarifier underflow into the lagoon (rather than the conventional practice of sludge haulage to waste disposal sites) improved system performance.

The process uses a combined single-stage activated sludge system with simultaneous carbon-oxidation and nitrification. High sludge age and high solids levels in the aeration cell of the pre-treatment unit contribute to high year-round nitrification. The low food to

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microorganism ratio encourages slow bacterial growth and endogenous respiration which minimizes excess sludge production. The concentration of nitrates released into the pond is high enough to prevent the reduction of sulphates to hydrogen sulphide, particularly in winter when anaerobic conditions are likely to occur. Denitrification takes place in the anaerobic sludge deposits in the pond where nitrates are converted to the gaseous form of nitrogen. Nitrogen gas is subsequently lost to the atmosphere. Controlled release of the waste activated sludge to the pond provides a carbon source for heterotrophic bacterial metabolism and thus promotes denitrification. The common practise of supplying an external carbon source for denitrification, e.g. methanol, is therefore not required.

If operated properly, this treatment system provides the dual benefits of 1) improved effluent quality in terms of hydrogen sulphide, ammonia, nitrates and bacteria and 2) reduced costs as little or no sludge haulage, or external carbon source for denitrification, is required. Sludge disposal costs represent a major operational expense for small sewage treatment facilities.

The results of the pilot studies were encouraging and it was decided by MOE to carry out a full-scale research project at the Sutton Water Pollution Control Plant (WPCP) to further field test and develop the process. The study was conducted by MOE, in cooperation with the Region Municipality of York, during the period from November, 1981 to May, 1984.

Description of the Sutton WPCP and Operational Conditions

The existing treatment facility at Sutton was chosen because it required minimal modification. The Sutton WPCP is of relatively simple design and construction (Table 1, Figure 1). A lined earthen basin with a mechanical surface aerator, followed by a clarifier with sludge collection and airlift sludge return, are the main components of the extended aeration pre-treatment unit. During full-scale testing of the process, the existing facilities (pre-treatment unit followed by waste stabilization pond) were maintained; but operational procedures, including the controlled release of sludge from the clarifier to the pond, were adjusted to improve system performance and reduce costs. No other pre-treatment (e.g. screening, grit removal) is provided at the facility. Clarifier effluent flows to the 6.5 ha (16 acre) polishing pond that discharges continuously to Lake Simcoe via the Black River.

Operating conditions following process conversion in November, 1981 are described in Table 2. Average raw sewage flow rates to the plant (0.26 MIGD) were 58% of design capacity (0.45 MIGD). Average hydraulic detention times were 44 hrs in the aeration tank, 7 hrs in the clarifier and 82 days in the pond. Average organic loadings were 184 kg BOD5/d (406 lbs BOD5/d); food to microorganism ratio (F/M) was generally less than 0.05 g(lbs)BOD5/g(lbs) MLVSS/d.

Results and Discussion

1. Carbon Oxidation and Nitrification

The combined carbon oxidation-nitrification stage of the process is characterized by high sludge age, high concentrations of biological solids and low food to microorganism ratio. The operational procedure of discharging waste activated sludge from the return sludge flow to the pond was used to control all three parameters. Sludge age, or solids retention time (SRT), was maintained at greater than the hydraulic detention time by recirculation of microorganisms through the sludge return flow. Because of the slow growth rate of nitrifying bacteria, a high SRT is necessary to retain an adequate population of nitrifiers. At the Sutton WPCP, wasting procedures (and the unintentional loss of solids in the clarifier effluent) resulted in a summer SRT range of 46-90 days and winter SRT range of 104-200 days.

The amount of waste activated sludge discharged varied during the year to compensate for the impact of temperature on process efficiency (Figure 2). The average monthly weight of solids released in the pond was 210 lbs SS/d. Due to increased bacterial respiration at high temperatures, sludge wastage was progressively increased at the start of the summer to lower aeration cell solids to levels which would not deplete the oxygen content of the wastewater and thereby inhibit nitrification. Wasting was gradually reduced at the approach of winter to build up solids levels (and the population of nitrifying bacteria) to compensate for the reduced activity of the nitrifiers at low temperatures. This is necessary as the nitrifier fraction of the total mass of biological solids, sustained by influent BOD5/TKN ratios of 4 to 5 measured during the study, has been estimated at only 5-6 % (1).

The plant was operated with the objective of maintaining mixed liquor suspended solids levels at 3000-4000 mg/L in summer and 4000-6000 mg/L in winter. Actual average MLSS levels were 3700-4817 mg/L in summer and 4154-5624 mg/L in winter (Figure 3). Volatile suspended solids ranged between 60 and 70% of the total suspended solids in the aeration cell. The excessive wasting that occurred in the summer of 1983, brought solids levels to less then 2400 mg/L in October, although the amount of sludge wasted was cut back in September and October. A further reduction in the suspended solids level down to 1790-2190 mg/L MLSS (1190-1390 mg/L MLVSS) for approximately two weeks in November resulted from a 5-day sewage bypass of the pre-treatment unit necessitated by mechanical repairs. During this period, the clarifier effluent ammonia levels rose to 15 mg/L. The data suggest that at these solids levels the population of nitrifiers were insufficient to support an adequate rate of nitrification. Slightly higher MLSS levels, 2380-2570 mg/L MLSS (1500-1610 mg/L MLVSS), measured in October and December of the same year, were not associated with any decline in aeration cell ammonia removal.

High solids levels in the aeration cell (overall average of 4594 mg/L MLSS) reduced the F/M ratios, during most of the study, to less than 0.05 g BOD5/g MLVSS/d. The low BOD5 loadings, in relation to the density of microorganisms, provided a starvation diet for the heterotrophic bacteria

as organic carbon required for cell synthesis and energy was kept in short supply. Rapid carbon oxidation under these conditions resulted in high BOD5 removal efficiencies (96-98%). In addition, the shortage of substrate slows the bacterial growth rate and encourages endogenous respiration whereby internal food reserves are utilized to support bacterial metabolism. Both are beneficial in wastewater treatment as they limit the production of excess sludge.

Average dissolved oxygen levels in the aeration tank were 1.6 mg/L in summer (range 0.-4.9 mg/L) and 5.9 mg/L in winter (range 2.1-9.8 mg/L). A decline in oxygen concentrations below 2 mg/L was observed during most measurements taken in May to September, 1982 and July to August, 1983. During these periods (including one occasion when measurements indicated a zero oxygen concentration), ammonia levels in the clarifier effluent remained below 0.5 mg/L. The average specific oxygen uptake rate (SUR) was 39% higher in summer (3.6 mg O2/L/hr/g MLVSS, range 1.9-5.7 mg O2/L/hr/g MLVSS) than in winter (2.6 mg O2/L/hr/g MLVSS, range 1.2-3.7 mg O2/L/hr/g MLVSS). The seasonal difference is the result of the elevated rate of bacterial metabolism at summer temperatures and the higher proportion of viable microorganisms in sludge with the short SRT's maintained in summer.

The levels of BOD5 and suspended solids in the clarifier effluent appeared to decline with increases in winter SRT but were little changed by variations in summer SRT (Figure 4). The improvement in secondary effluent quality with respect to BOD5 and suspended solids may be related to the observed reduction in the sludge volume index (SVI) with increases in winter season SRT (Figure 5).

Ammonia removal in the aeration cell (from average influent levels of 23 mg/L) remained high in winter during which time ammonia concentrations in the clarifier effluent infrequently exceeded 0.5 mg/L (Table 3). Elevated ammonia levels above 1 mg/L in the clarifier effluent were generally associated with equipment breakdowns that adversely affected nitrification. Nitrification rates were affected by malfunction of the ferric chloride feed line in January, 1982 (during which time application rates rose above 40 ppm) causing acidification of the mixed liquor and by depressed solids levels resulting from a combination of excess wasting plus sewage bypass which occurred in November, 1983. The impact of these elevated clarifier effluent ammonia levels on the final pond effluent were limited as the average pond effluent levels were as low as 1.1 mg/L (max 2.2 mg/L) in the first winter (1981/82), 0.42 mg/L (max 1.1 mg/L) in the second winter (1982/83) and 2.0 mg/L (max 3.8 mg/L) in the third winter (1983/84) when ammonia levels entering the pond rose to 15 mg/L for 2-3 weeks. Although studies indicate a severe reduction in the rate of nitrification at temperatures less than 10 oC (2), a pattern of reduced ammonia removal at these temperatures was not apparent in the data.

2. Denitrification and Waste Activated Sludge Reduction

Measurements conducted in February, 1986 indicate that the waste activated sludge discharged to the pond accumulated in a circular sludge pile, diameter 46 m (150 ft), beginning at the clarifier effluent discharge pipe in the pond. The depth of the sludge deposit ranged from 45 cm (18 inches) near the outlet to 8-10 cm (3-4 inches) at the perimeter. Assuming an average depth of 15 cm (6 inches), the total sludge volume accumulated since November, 1981 is estimated at approximately 785 m3 (27,730 ft3).

Sludge wasted during the first winter (1981/82) contained an average of 1.1% total solids (based on weight measurements, mg/kg) with volatile substances comprising 69% of the total solids. Using these values as starting points, the digestion and reduction of the bottom sludge (within 20 ft of the waste outlet to the pond) was monitored during four years following process conversion and is plotted in Figure 6. The data indicate an increase in percent total solids and a decrease in the volatile portion to asymtotes of 41% total solids and 18% organic matter reached in 2 1/2 years. Sludge reduction is accomplished under anaerobic conditions by heterotrophic bacteria capable of nitrate respiration in the absence of oxygen. The denitrifying bacteria obtain their carbon and energy from endogenous respiration (3).

The process of denitrification removed 71-95% of the nitrate released to the pond in summer and 47-71% during the winter period (Table 4). Although the denitrifying bacteria are not as sensitive to low temperatures as the nitrifiers (denitrification is reported to occur over a temperature range of 0-50 oC), low temperatures have been shown to reduce the rate of denitrification (4).

3. Pond Effluent Quality

The range of seasonal average BOD5 and suspended solids levels in the pond effluent was 3-12 mg/L and 3-7 mg/L, respectively (Table 5). Monthly average suspended solids levels were less than 19 mg/L (generally less than 10 mg/L) whereas BOD5 levels were less than 13 (generally less than 6 mg/L) (Figure 7). The pattern of monthly fluctuations in the two parameters were similar, with most or all of the reduction occurring prior to the release of the clarifier effluent into the pond.

Seasonal average ammonia and total Kjeldahl nitrogen (TKN) concentrations were reduced to 0.3-2.5 mg/L and 1.8-4.0, respectively (Table 6). Reductions in summer and winter ranged between 89 and 98%. Ammonia and TKN levels in the pond effluent were often higher than levels measured in the clarifier effluent, which were as low as 0.01 mg/L in summer. Increases in ammonia concentrations within the pond were higher in summer (100 to 450%) than in winter (-20 to 47%) due to the generation of ammonia from organic decomposition in the pond and the very low levels of ammonia achieved in the clarifier effluent in summer.

Monthly average ammonia concentrations in the pond effluent were often less than 0.5 mg/L and generally less than 1.0 mg/L (Figure 8). Levels exceeding 2.0 mg/L occurred in the winter of 1983 as a probable direct

result of a raw sewage bypass to the lagoon and indirectly through a failure to adequately nitrify when solids levels fell below 2300 mg/L MLSS. Monthly average nitrate concentrations in the pond effluent (Figure 8) were generally less than 5.0 mg/L with increases to annual peak levels of 6.7-10.8 mg/L toward the end of winter (i.e. February, March or April).

No hydrogen sulphide was detected in the pond effluent during the study period. Although nitrate levels in the pond were generally high enough to satisfy the oxygen requirement of the heterotrophic bacteria, when nitrate levels declined in the summer of 1983, there was sufficient oxygen generated by algae to maintain aerobic metabolism in the pond.

Continuous dosages of ferric chloride were applied to the wastewater, following discharge from the aeration tank and prior to entry into the clarifier, for phosphorus precipitation/coagulation. The initial average application rate of 7 mg/L Fe +3 during the first winter (1981/82) was raised to 15 mg/L Fe +3 during the second winter (1982/83) (Table 7) and maintained at elevated levels for the duration of the study. During the first summer (1982), monthly average total and soluble phosphorus levels in the pond effluent rose above 1.0 mg/L for several months (Figure 9). These levels reflected the low Fe +3 additions (monthly average dosages as low as 5 mg/L) during the first year of the study. Total phosphorus levels in the pond effluent increased slightly in the summer months, but as a result of increased ferric chloride dosages, were consistently less than 1.0 mg/L and usually less than 0.5 mg/L. Soluble phosphorus concentrations showed the same pattern of variation.

Fecal bacteria reductions in the pond ranged between 95.1 and 99.8% (Table 8). Elevated pond effluent bacteria levels were observed at the end of February and in March in some years. These may have resulted from short-circuiting under the ice or from the impact of flushing on pond retention times during the period of ice melt. Bacterial levels in samples taken along a transect from the clarifier effluent outlet to the pond outlet indicate that most of the reduction of fecal bacteria takes place near the influent end of the pond.

Conclusion

Because of the hazards to the biological integrity of limited receiving waters from year-round or seasonal discharge of waste stabilization pond effluents, there is a widespread potential application of this process in the design of new sewage treatment facilities and upgrading of existing facilities. The data presented here illustrate the high level of treatment that can be achieved by the process when operated under prescribed conditions. Due to the need for more information on process design and operation, the Water and Wastewater Management Section of the Water Resources Branch is presently supporting a study of the Colborne WPCP, a sewage treatment facility which underwent process conversion in 1983. The objectives of the study are to 1) develop operational guidelines for achieving optimal system performance, 2) establish specifications for pond sizing and hydraulic detention times to achieve denitrification and at the same time reduce the frequency of sludge

disposal, 3) determine minimum effluent total phosphorus achievable (and associated chemical dosages and costs) and 4) investigate the dynamics of fecal bacteria reductions in the pond.

References

1. U.S.EPA. 1975. Process Design Manual for Nitrogen Control. Office of Technology Transfer, Washington, D.C.

2. Shammas, N.K. 1986. Interactions of temperature, pH and biomass on the nitrification process. JWPCF 58(1): 52-59.

3. Sutton, P.M., B.E. Jank, B.A. Monaghan and K.L. Murphy. 1979. Single stage sludge nitrogen removal systems. Canada-Ontario Agreement on Great Lakes Water Quality.

4. Sutton, P.M. and B.E. Jank. 1976. Design considerations for biological denitrification systems.IN: High Quality Effluents Seminar. Conference Proceedings No.3. Canada-Ontario Agreement on Great Lakes Water Quality.

Table 1. Design Criteria for the Sutton WPCP

ADWF	*0.45 MIGD*
Aeration Tank Volume	*0.45 MIG*
Aeration Detention	*24 hrs at ADWF*
Clarifier Volume	*0.07 MIG*
Clarifier Detention	*4.3 hrs at ADWF*
Clarifier Upflow Rate	*17 m3/m2/d (349 gpd/ft2) at ADWF*
Sludge Return Rate	*up to 125% of ADWF*
Pond Area	*6.5 ha (16 acres), 1.8 m (6 ft) max. depth*
Pond Volume	*20 MIG at 1.5 m (5 ft) normal op. depth*
Pond Detention	*44 days at ADWF*

Table 2. Sutton WPCP Operating Conditions (Dec.,1981-Oct.,1985)

Parameter [units]		Average *	Range
Hydraulic Loading [MIGD]		0.26	0.16-0.64
Influent BOD5 [mg/L]		167	70-352
Organic Loading [kg BOD5/d]		184	62-588
[lbs BOD5/d]		406	137-1296
F/M [g(lb) BOD5/g(lb) MLVSS/d]		0.03	0.01-0.18
Volume Loading [kg BOD5/m3/d]		0.10	0.03-0.29
[lbs BOD5/103ft3/d]		5.8	1.9-18.0
Aeration Cell HDT [hrs]		44	19-66
Aeration Cell MLSS [mg/L]		4594	summer: 3700-4817 winter: 4154-5624
Sludge Age [days]	summer:	67	46-90
	winter:	147	104-200
Sludge Volume Index		150	82-200
Aeration DO [mg/L]	summer:	1.6	0.-4.9
	winter:	5.9	2.1-9.8
Aeration Temperature [oC]	summer:	16.3	9.5-22.0
	winter:	5.1	0.5-11.5
Clarifier HDT [hrs]		7	3-11
Clarifier Upflow Rate [m3/m2/d]		9.8	
[gpd/ft2]		200	
Sludge Return Rate [% of avg. flow]			100-150
Pond HDT [days]		82	monthly avg: 59-113
WAS Discharge To Pond [lbs SS/d]		210	monthly avg: 0-675

* Mean of winter (Nov-Apr) and summer (May-Oct) averages unless specified.

Table 3. Ammonia Concentrations in the Clarifier Effluent (mg/L)

Season	Avg.Conc.	Range	S.D.	% Samples (or=0.5 mg/L	No. of Samples
Winter 81/82*	0.75	0.04-7.0	1.6	82	44
Summer 82	0.19	0.01-0.50	0.1	100	29
Winter 82/83	0.36	0.01-4.7	0.9	92	24
Summer 83	0.22	0.05-0.90	0.2	94	16
Winter 83/84**	2.50	0.10-15.7	5.0	63	16

* *Excessive dosages of FeCl3 caused acidification of ML in Jan/82.*
** *Depressed MLSS from excess wasting and sewage bypass in Nov/83.*

Table 4. Nitrate Removal in Pond

Season	Clarifier Effluent Avg. Conc.(mg/L)	Pond Effluent Avg. Conc.(mg/L)	% Reduction
Winter 81/82	16.3	6.6	60
Summer 82	18.8	1.0	95
Winter 82/83	19.0	5.5	71
Summer 83	16.3	2.7	83
Winter 83/84	15.3	7.9	48
Summer 84	13.9	4.1	71
Winter 84/85	16.0	8.5	47
Summer 85	8.2	2.3	72

Table 5. Average BOD-5 and Suspended Solids Concentrations (mg/L)

Season	BOD-5				Suspended Solids			
	Raw Sew.	Clar. Eff.	Pond Eff.	% Reduct.	Raw Sew.	Clar. Eff.	Pond Eff.	% Reduct.
Winter 81/82	146	15	6	96	208	38	10	95
Summer 82	172	4	4	98	206	9	8	96
Winter 82/83	163	8	7	96	124	13	12	90
Summer 83	124	4	5	96	131	7	7	95
Winter 83/84	-	6	6	-	-	9	9	-
Summer 84	284	3	6	98	235	6	8	97
Winter 84/85	158	5	3	98	139	11	5	96
Summer 85	125	6	3	98	135	7	3	98

Table 6. Average Ammonia-N and TKN Concentrations (mg/L)

Season	Ammonia-N				TKN			
	Raw Sew.	Clar. Eff.	Pond Eff.	% Reduct.	Raw Sew.	Clar. Eff.	Pond Eff.	% Reduct.
Winter 81/82	19	0.8	1.1	94	31	3.9	3.3	89
Summer 82	22	0.2	1.1	95	33	1.8	3.1	91
Winter 82/83	23	0.4	0.4	98	34	2.1	2.5	93
Summer 83	21	0.2	0.6	97	31	1.5	2.6	92
Winter 83/84	-	2.5	2.0	-	-	3.7	4.7	-
Summer 84	27	0.2	0.3	99	36	1.0	2.0	94
Winter 84/85	26	1.7	2.5	90	40	2.7	4.0	90
Summer 85	23	0.1	0.5	98	30	1.0	1.8	94

Table 7. Average Total and Soluble Phosphorus Concentrations (mg/L)

Season	Total Phosphorus				Soluble Phosphorus				FeCl3 Dosage (mg/L Fe+3)
	Raw Sew	Clar. Eff.	Pond Eff.	% Reduct.	Raw Sew.	Clar. Eff.	Pond Eff.	% Reduct.	
W 81/82	5.6	1.5	0.4	93	3.3	0.5	0.3	91	7
S 82	7.0	1.4	1.5	79	4.2	1.0	1.2	71	10
W 82/83	6.8	0.5	0.3	96	3.9	0.1	0.1	97	15
S 83	5.4	0.4	0.4	93	3.3	0.2	0.2	95	14
W 83/84	-	0.3	0.2	-	-	0.1	0.03	-	12
S 84	7.4	0.2	0.3	96	-	0.1	0.04	-	12
W 84/85	7.4	0.3	0.2	98	-	-	-	-	9
S 85	6.2	0.6	0.4	94	-	-	-	-	8

Values are summer(S) and winter(W) averages.

Table 8. Fecal Bacteria Reductions in the Pond

Total Coliform

Season	Clarifier Effluent	Pond Effluent	% Reductions
Winter 81/82	112,293	795	95.3
Summer 82	70,000	320	98.1
Winter 82/83	61,747	96	99.8

Fecal Coliform

	Clarifier Effluent	Pond Effluent	% Reductions
Winter 81/82	10,345	91	93.1
Summer 82	22,000	10	99.5
Winter 82/83	15,986	15	99.8

Fecal Streptococcus

	Clarifier Effluent	Pond Effluent	% Reductions
Winter 81/82	3,380	30	97.3
Summer 82	1,200	10	95.1
Winter 82/83	1,334	11	99.1

Bacteria are geometric mean levels in number per 100 ml.

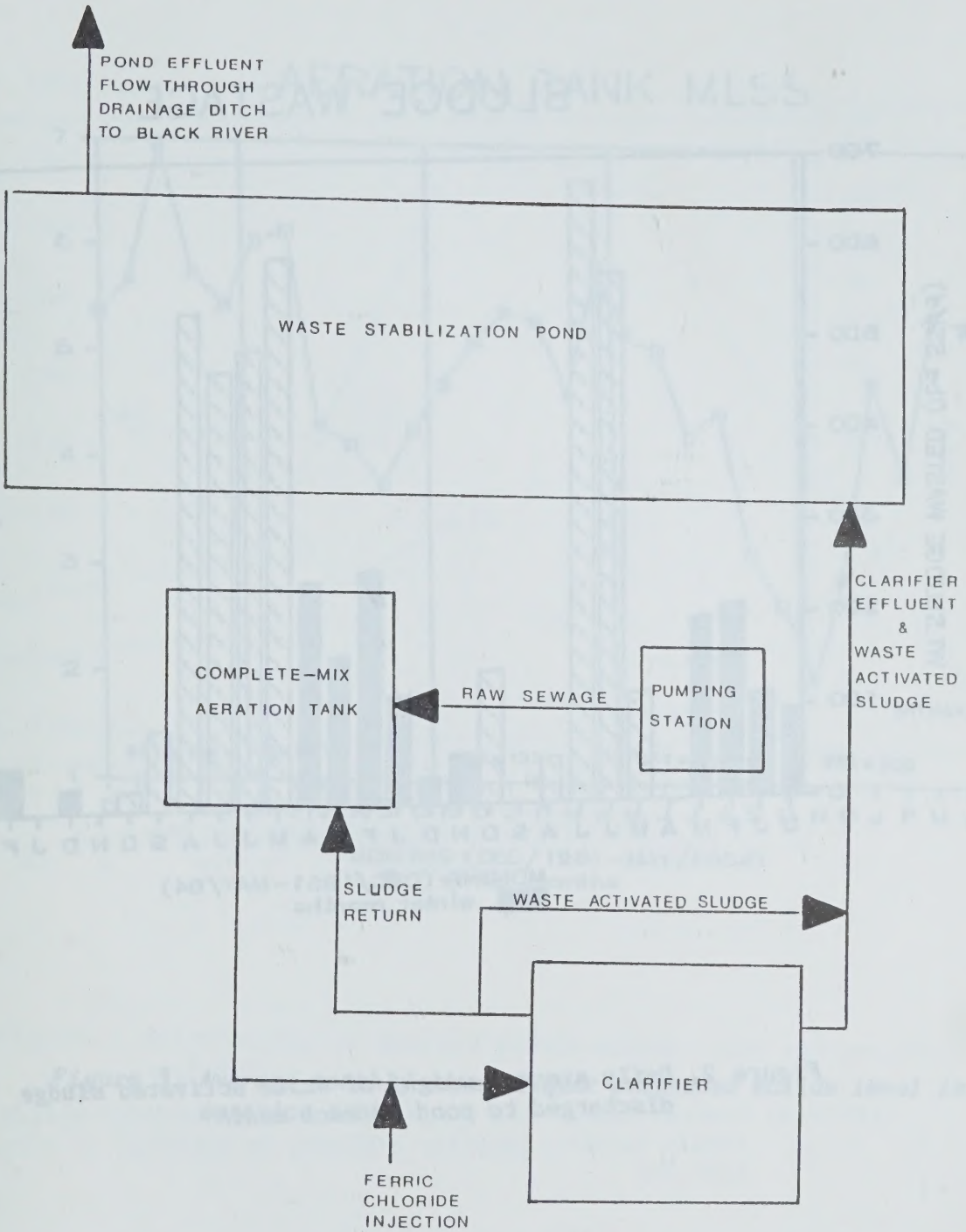


Figure 1. Diagram of the Sutton WPCP

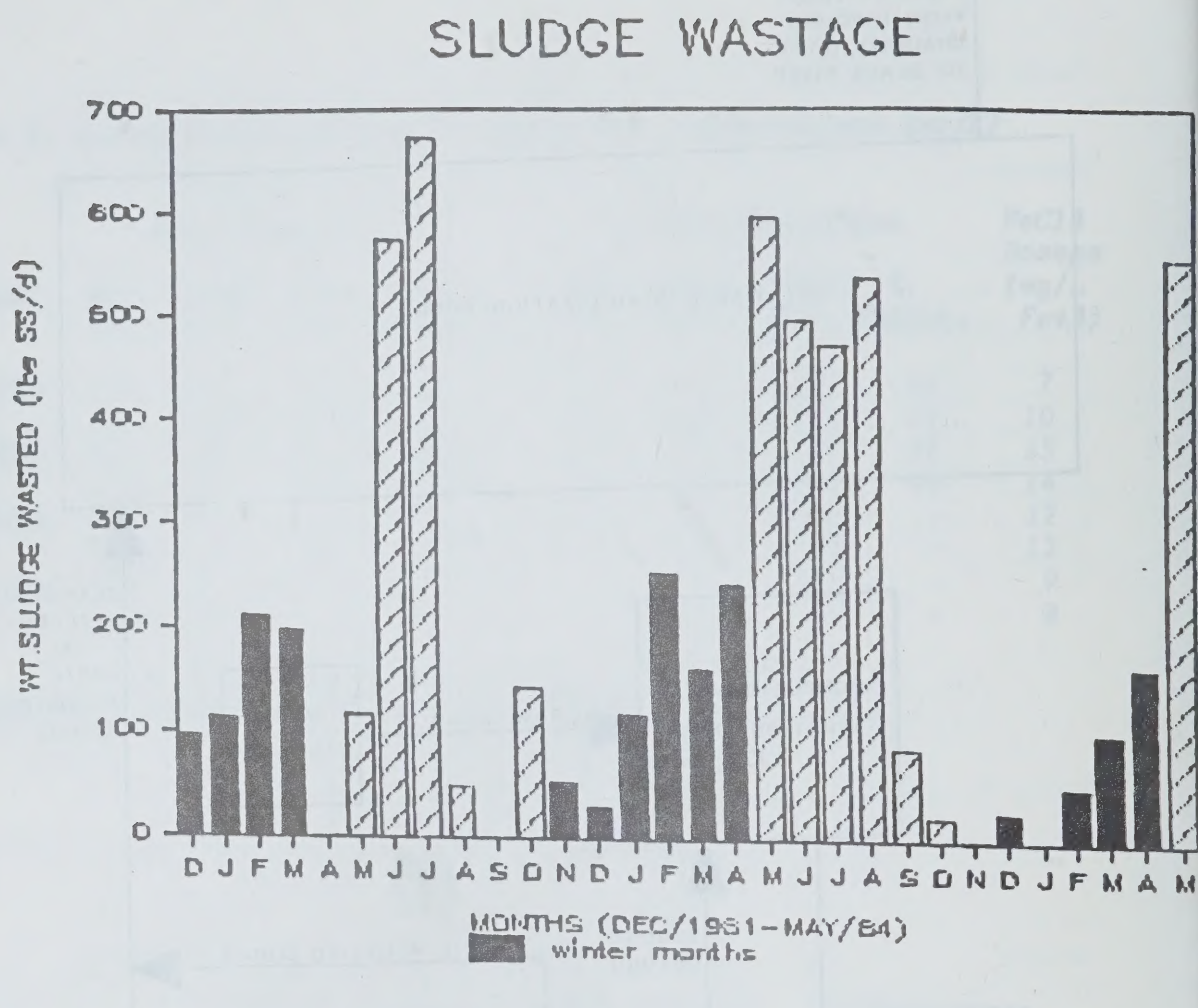


Figure 2. Daily average weight of waste activated sludge discharged to pond in each month

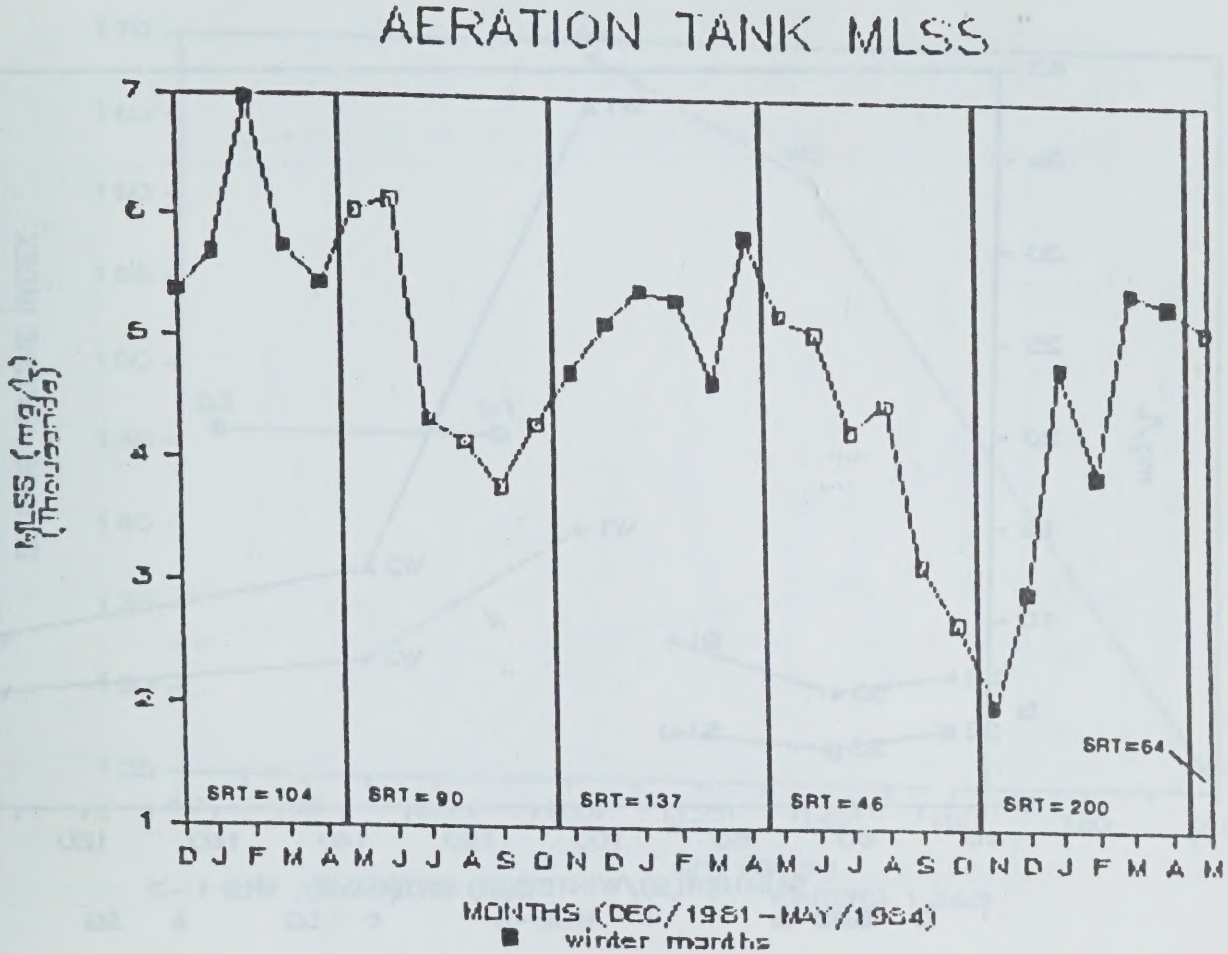


Figure 3. Average monthly mixed liquor suspended solids level in aeration tank

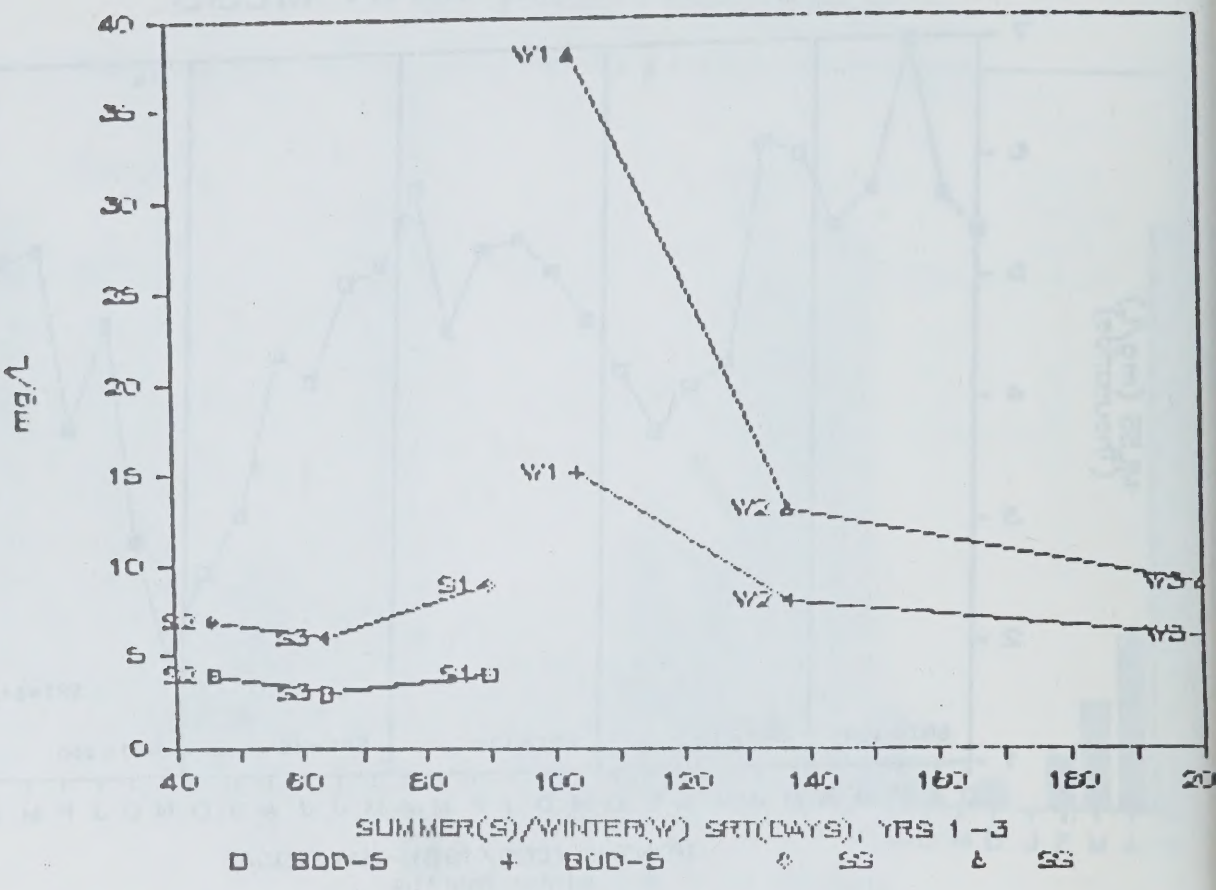


Figure 4. Average summer and winter BOD-5 and suspended solids levels in the clarifier effluent in relation to aeration tank SRT

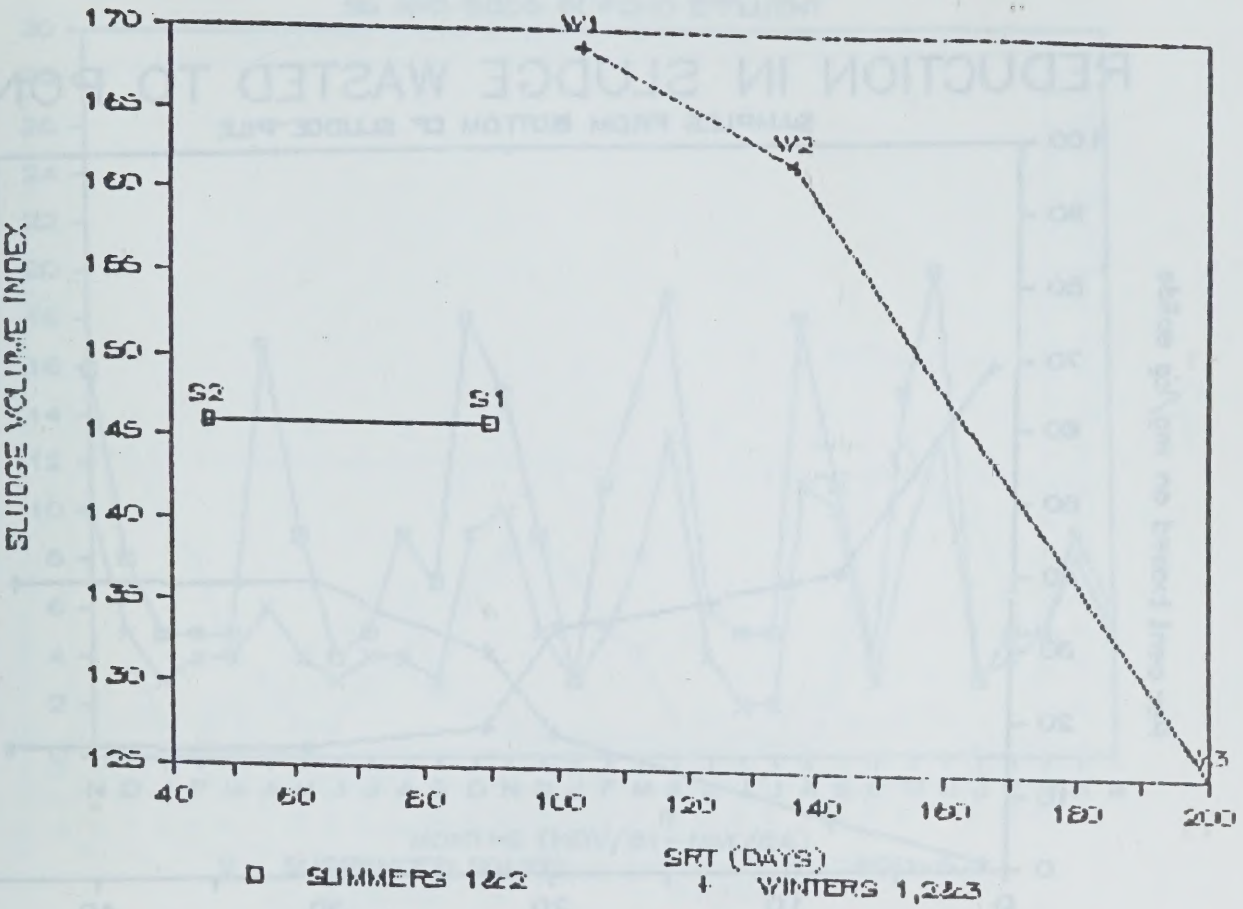


Figure 5. Relationship of average sludge volume index and aeration tank SRT

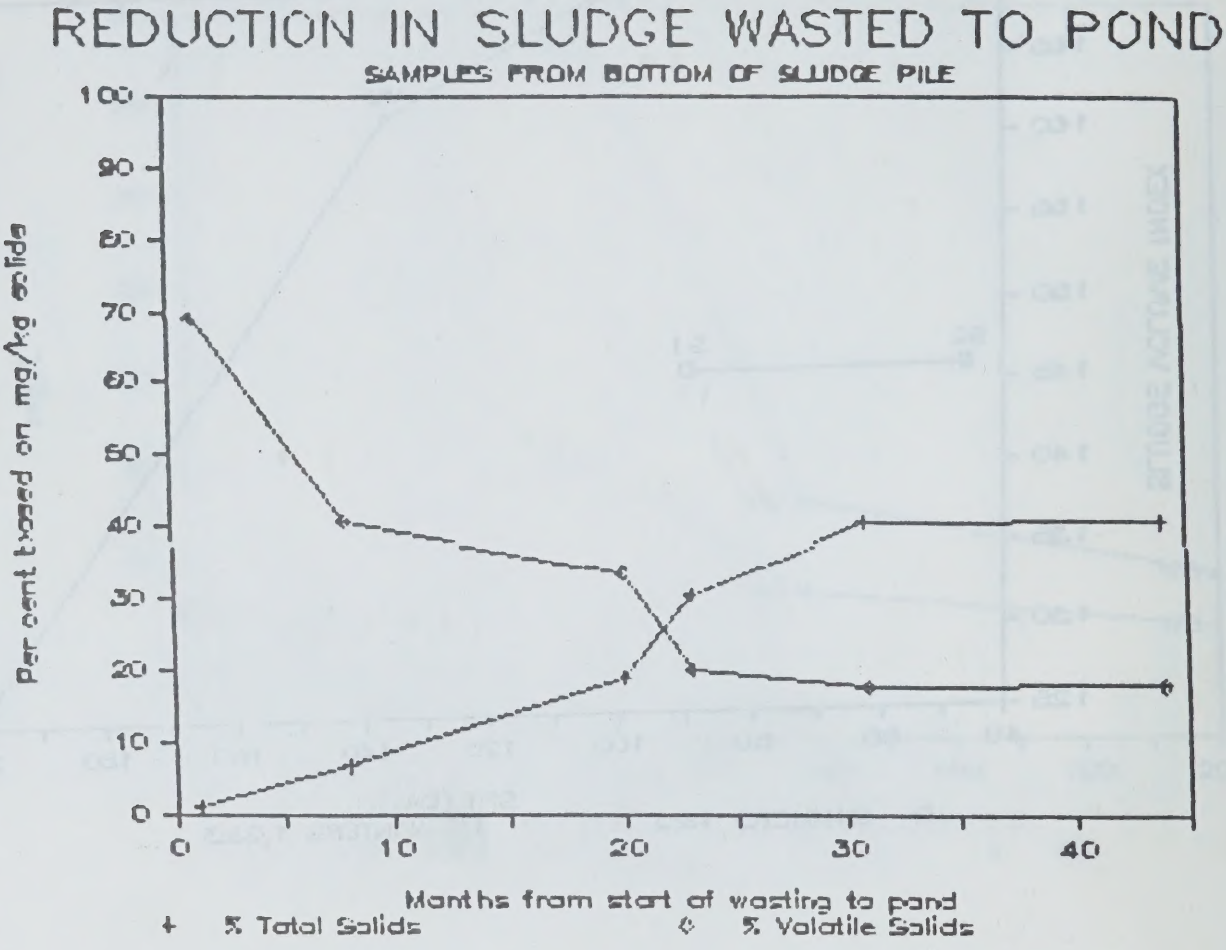


Figure 6. Changes in percent total solids and the volatile percentage of the total solids in the bottom of the waste activated sludge deposits in the pond

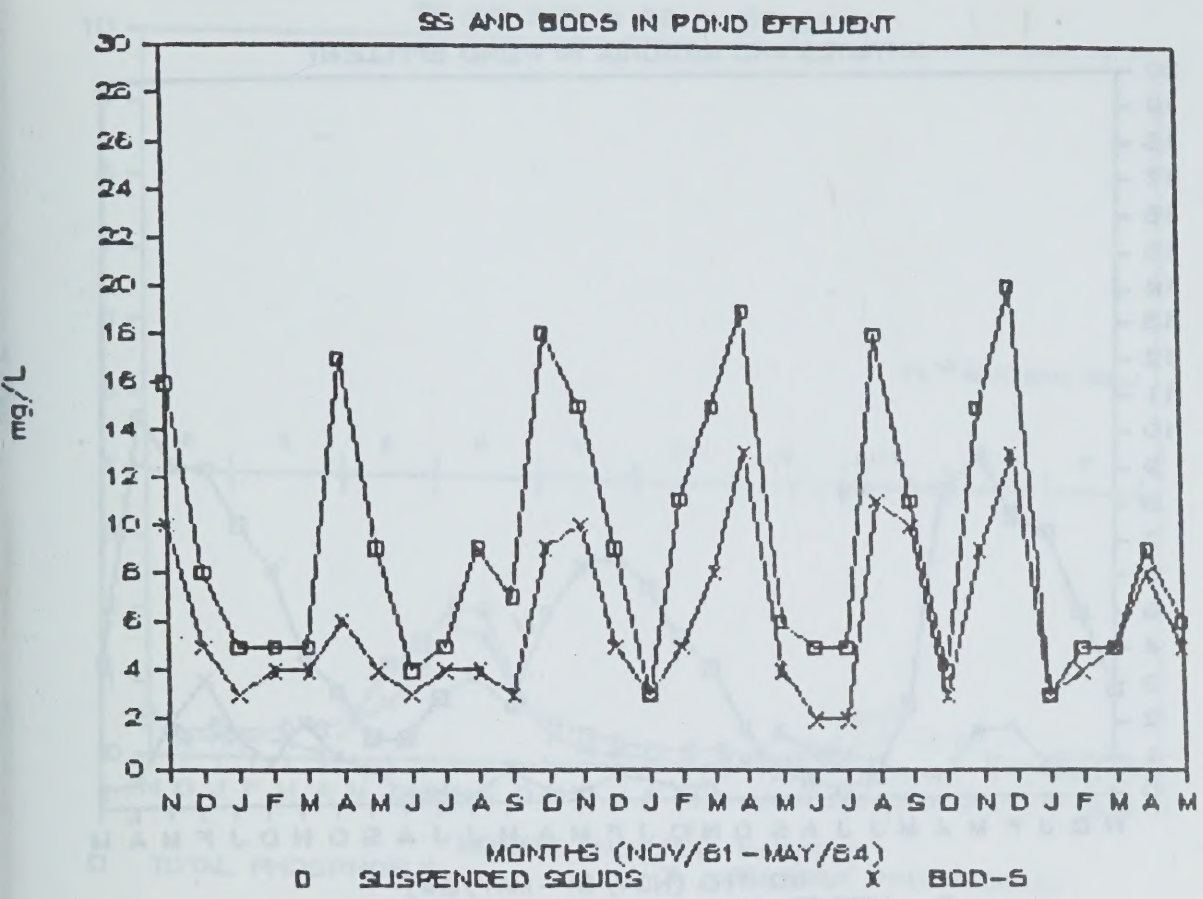


Figure 7. Monthly average suspended solids and BOD-5 concentrations in the pond effluent

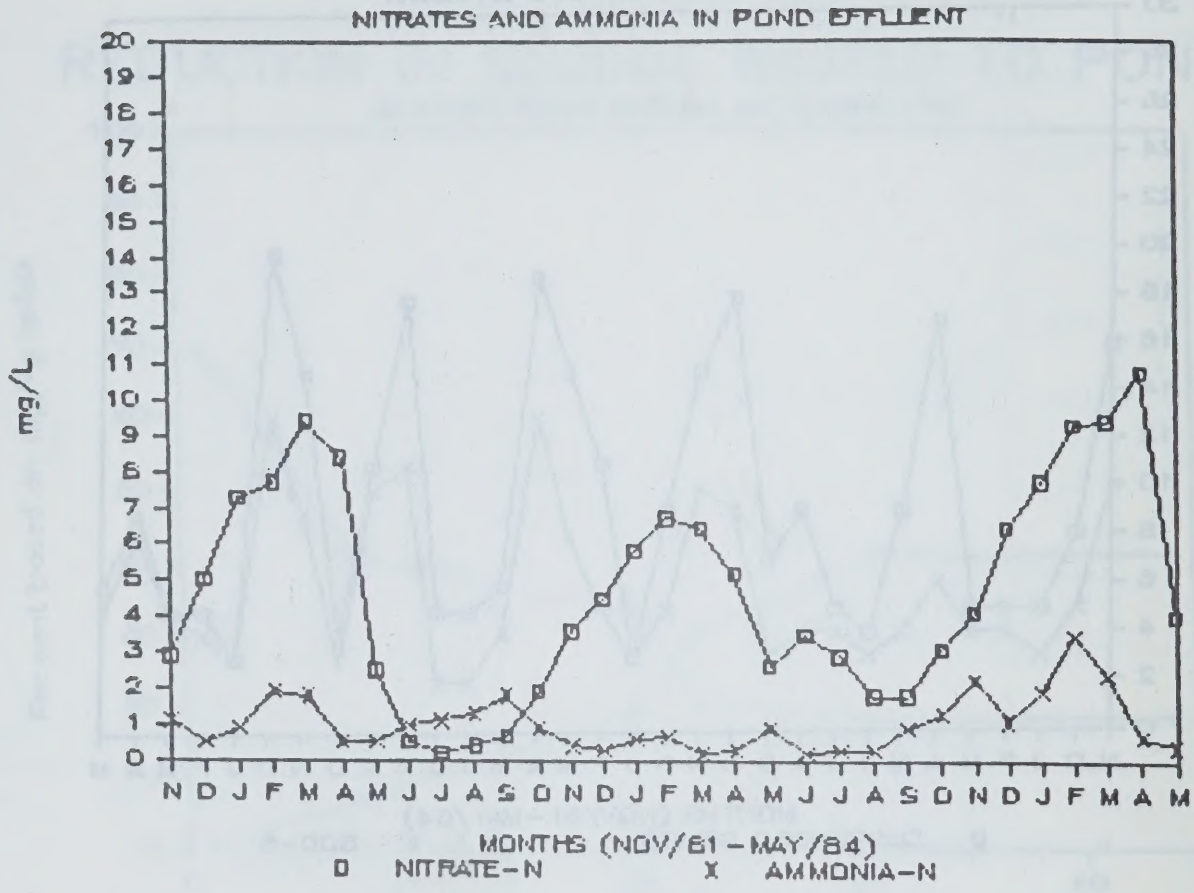


Figure 8. Monthly average nitrate and ammonia concentrations in the pond effluent

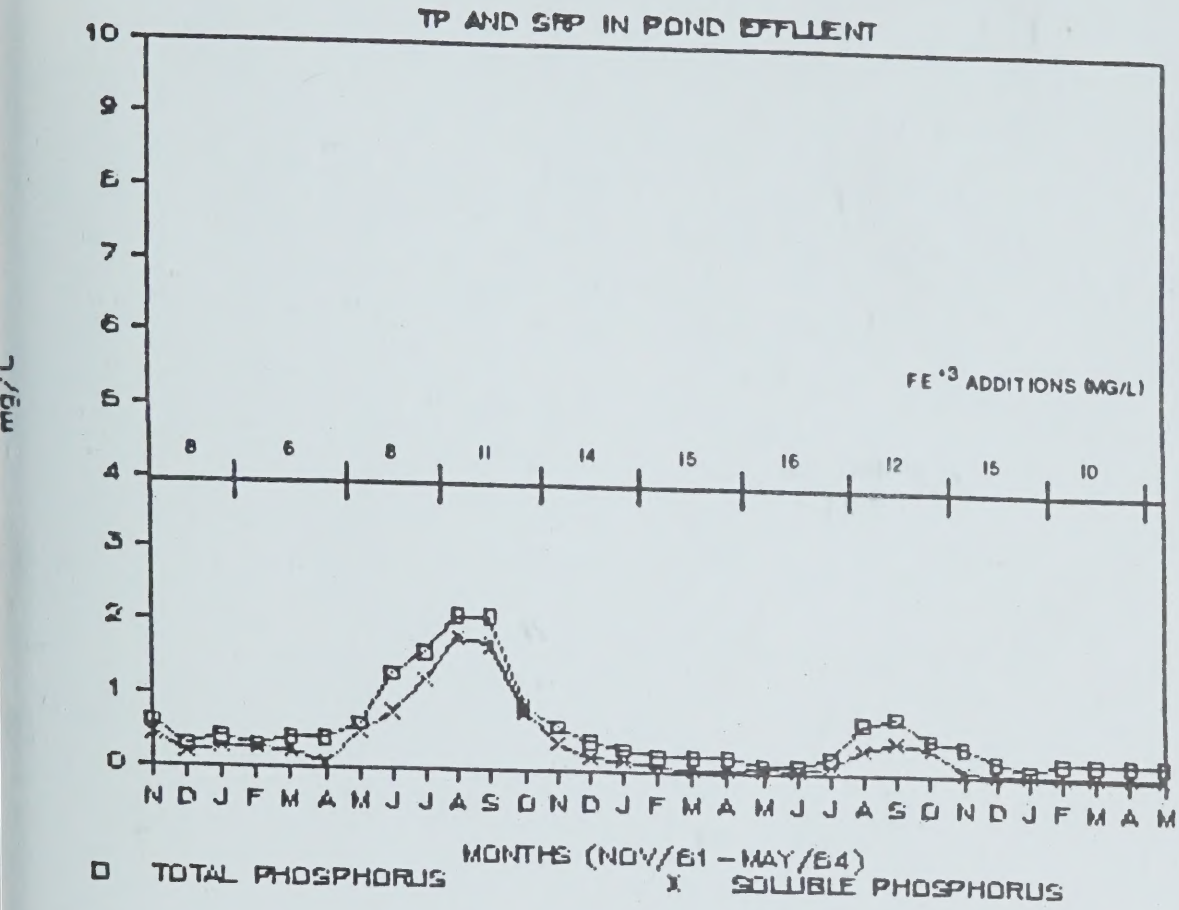
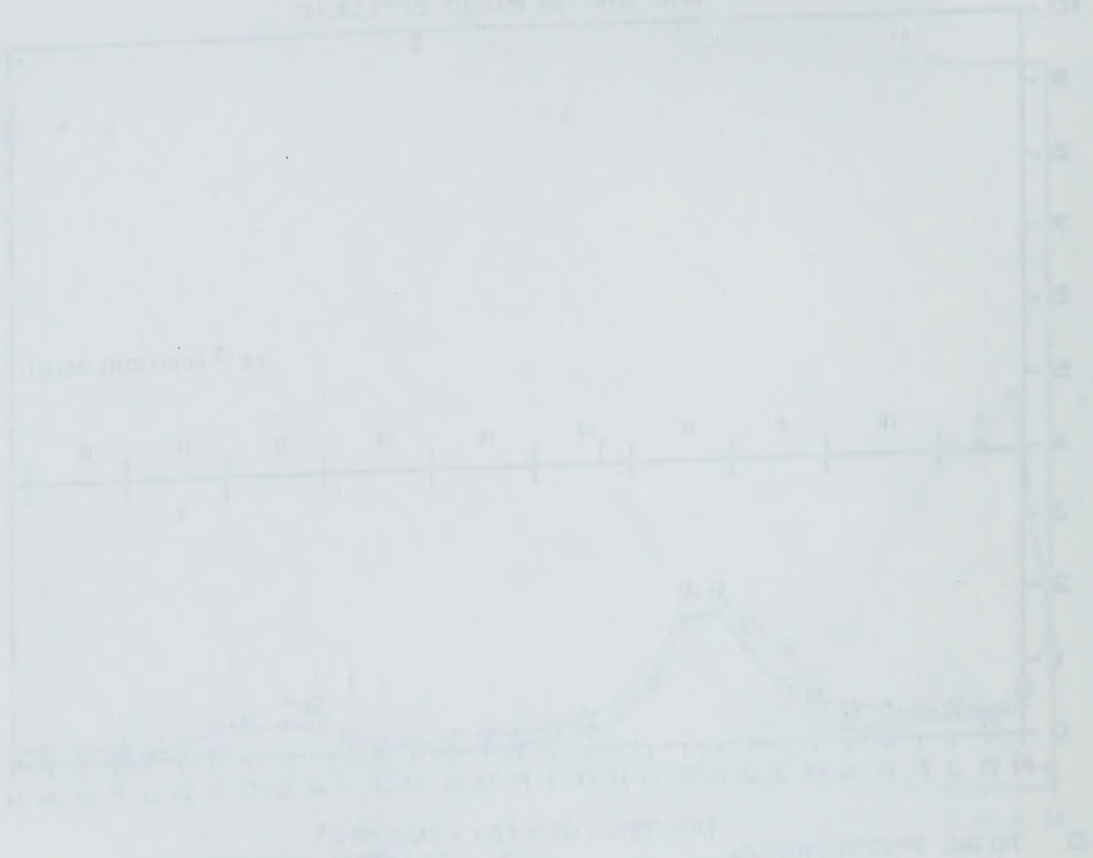


Figure 9. Monthly average total and soluble phosphorus concentrations in the pond effluent



THE EFFECT OF APPLIED DIRECT CURRENT ON BIOLOGICAL PHOSPHORUS UPTAKE

by

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ABSTRACT

It was hypothesized that the introduction of direct current into mixed liquor might change the energy level of the bacterial cells, increasing the activity or the percentage of bacteria capable of accumulating excess phosphorus.

A direct current with 6 V and 200 mA was applied for one minute in every five. The direct current was introduced at the anaerobic step of the Phoredox type laboratory scale continuous flow system. Graphite electrodes were used to introduce the current. To separate the effect of the electric current from that caused by other environmental factors a control system was run in parallel using the same operational parameters.

As a result of the intermittent application of direct current, the rate of phosphorus accumulation in activated sludge increased by 30% during the start-up period of the biological system and by 11% during steady state condition. The applied direct current changed the movement and concentration of potassium ions. The pH of the mixed liquor dropped to 4 as soon as the direct current was introduced and returned to the initial pH 6 value when the current flow ceased. The number of viable cells decreased by 2 to 5 times as a result of electrification. The specific cell activity was higher in the electrified than in the control system. The reduction observed in viable cell count did not influenc the overall organic carbon removal efficiency nor the nitrification, denitrification processes.

A hypothetical model was developed based on the observed results. The model summarises the effects of direct current on biological phosphorus uptake.

LITERATURE REVIEW

Enhanced biological phosphorus removal in activated sludge treatment plants can be achived by using consecutive anaerobic and aerobic treatment steps (1).

All operational and experimental results emphasize the role of the anaerobic step as the prerequisite to increase the population of aerobic bacteria capable of accumulating excess phosphorus (1),(2),(3).

Under anaerobic conditions, bacteria which are capable of accumulating excess phosphorus, remove the easily biodegradable organic substrate from the wastewater causing a significant drop in COD concentration. At the same time the phosphorus concentration of the filtrate increases (4).

Under anaerobic conditions the hydrogen transfer mechanism in bacterial cells is interrupted because of the absence of a hydrogen acceptor. In this situation only 5% of the total energy stored in the food can be utilized which is not enough for aerobic bacteria. Certain types, however, are able to store energy in the form of polyphosphate and under anaerobic condition they mobilize it to synthetize ATP. During this procedure phosphorus is released from the bacterial cells. The extra energy gained from polyphosphate is then used to carry on essential cell functions and to accumulate excess food stored as polyhydroxybutyrate (PHB). PHB is primarily synthetized from easily biodegradable COD such as sugars and short chained fatty acids (5),(6).

Under aerobic condition the stored PHB is the extra energy source used to build up the polyphosphate content of the cell and to gain energy if the external food sources are deplated (5).

By building up the polyphosphate content of the cell, phosphorus is removed from the wastewater at a higher rate than its rate of release during the anaerobic step (1),(3).

The easily biodegradable COD is the carbon source required for denitrification which step is necessary to establish the anaerobic environment needed for the biological phosphorus removal. However, the same carbon source is used to build up the PHB content of the cell (7),(8).

The relationship between easily biodegradable COD and total COD as well as the TKN/COD ratio in the influent will determine, whether there is enough easily biodegradable COD to complete the denitrification and to build up the required PHB level in the cells (5).

The composition of wastewater can be improved by the addition of short chained fatty acids originating from the supernatent of anaerobic digesters (9).

The energy transfer processes of the cell takes place at the cell membrane. The membrane acts as a selective filter regulating the movement of nutrients into and out of the cell. To ensure adequate membrane and cell function the electrochemical potential

of the membrane should be constant (10),(11).

The electrochemical potential of the membrane is determined by the hydrogen and potassium ion gradients. The proton gradient is used to co-transport important nutrients such as sugars and amino acids (10),(11),(12).

The cell in order to regulate its internal pH and to keep the electrochemical potential of the membrane constant, has to be able to control proton and cation movements by using the ATPase, K/H and Na/H transport systems (13),(14).

The ATPase, K^+/H^+ systems require energy obtained by the hydrolysis of ATP to remove H^+ from the cell. The Na^+/H^+ antiport is possibly more active in the removal of Na^+ using ATP to expel it and take up H^+ (15),(16).

The cell can temporarily gain energy in the form of ATP by reversing the ion transport of the ATPase, K^+/H^+, Na^+/H^+ transport systems (16).

Electric current may change the pH of an aqueous solution if its buffering capacity is not large enough (17),(18).

Externally applied electrical energy can be utilized by the bacterial membrane to produce ATP, thus the membrane can transfer electrical energy into chemical form and vica versa (19).

It was observed that at a critical transmembrane potential, caused by external electrical energy, local instabilities or the formation of hydrophilic pores may occure thereby changing the permeability of the membrane. After the electric field is removed a resealing process takes place (20).

HYPOTHESIS FOR THE EFFECT OF EXTERNAL ELECTRICAL ENERGY ON BIOLOGICAL PHOSPHORUS UPTAKE

The literature review summarized in the previous section demonstrates the importance of the electrochemical potential in the membrane transport and energy transfer processes of the bacterial cell. The fact that externally applied electrical energy can be transfered to chemical energy via the cell membrane suggests that excess biological phosphorus uptake could be stimulated by using electric current.

It was hypothesized that the application of direct current into mixed liquor might change the energy level of the bacterial cells, increasing the activity or the percentage of bacteria capable of accumulating excess phosphorus.

It was proposed to apply the direct current at the anaerobic step of the biological system, because this is the place where the bacterial cells have limited access to energy, thus any small change in energy level could be significant.

It was thought, that the overall effect of direct current on the bacterial cell metabolism might be influenced by the amount of electrical energy introduced and by the intermittent nature of the current flow.

EXPERIMENTAL SYSTEM

It was decided to use a Phoredox type system which may have a high phosphorus as well as nitrogen and COD removal efficiency.

The long hydraulic and solids retention time applied in the system provide buffering capacity which permits the maintenance of steady state easily.

The system consisted of an anaerobic, aerobic and a final settling tank with sludge recirculation. To separate the effects of electrical current from that caused by other environmental factors a control system was introduced in parallel using the same operational parameters. The volume of tanks and the operational parameters of the experimental system are shown in Fig.1.

The electrodes used to apply the electric current were mounted in the anaerobic tank of the electrified system. The material chosen for the electrodes was graphite. The electrodes had a surface area of 180 cm^2 and were placed 5.5 cm apart.

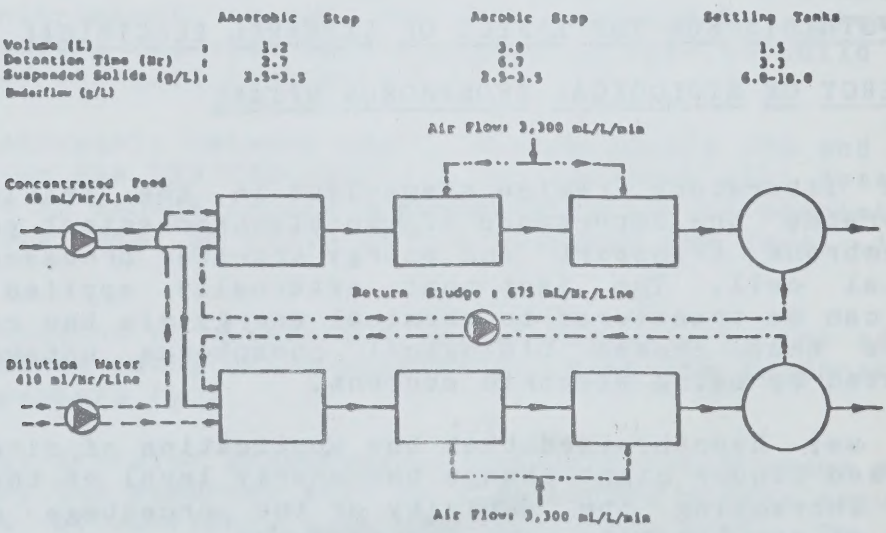

Fig.1 Flow Diagram of Experimental System

The direct current introduced to the anaerobic reactor was regulated by a timer. It was decided to use one minute flow of direct current in a five minute period applying 6 V and 200 mA.

The composition of wastewater has an important effect on the efficiency of the phosphorus removal system. Synthetic wastewater was used in order that the composition could be maintained constant. Three different organic carbon sources (Dextrose, Beef extract, Peptone) were added to ensure a well balanced diet for the biological system. The flow of organic carbon substrates was controlled to provide an inflow of 500 mg COD/L concentration. The inorganic ions, such as nitrogen, phosphorus, sodium, manganese, potassium, and chloride required for biological metabolism, were added in a constant quantity to the synthetic wastewater. The phosphorus concentration of the inflow was maintained at 8.5 mg/L throughout the study (21).

ANALYTICAL METHODS

The tests carried out included the measurement of phosphorus, potassium, sodium, pH, Chemical Oxygen Demand (COD), different forms of nitrogen and solids concentration, and additional tests such as cell activity and plate count. The tests were carried out on samples taken from the effluent of each tank of both systems.

The total phosphorus content of the filtrate and the mixed liquor were measured to obtain the phosphorus content of the sludge expressed as a percentage of volatile solids. The samples were pre-digested and analysed on a Technicon auto -analyser.

The potassium and sodium concentration of the filtrates were measured and expressed as a percentage of the volatile suspended solids content of the sample. The concentration of these elements were measured with an Atomic Absorption Spectrophotometer using the Emission operational mode.

The Chemical Oxygen Demand (COD), Total Kjeldahl Nitrogen (TKN), ammonium nitrogen, and the concentration of total nitrogen oxide (NO_x) were determined in accordance with "Standard Methods" (22).

The activity of the sludge was measured using Triphenyltetrazolium Chloride (TTC). The TTC serves as an artificial hydrogen acceptor measuring the dehydrogenese activity of the cell (23).

The activity test was supplementad with plate count using the smear technique described by the "Standard Methods", to determine the number of viable cells present in sludge (22).

RESULTS

Batch scale tests showed that the application of direct current does not have any immediate effect on phosphorus release or uptake. These results underlined the need for tests conducted on continuous flow systems, providing enough time for the establishment of a bacterial population capable of accumulating excess phosphorus and developing a measurable response to the electrical stimulus.

It took about eight weeks for the continuous flow system to increase the bacterial population capable of storing excess phosphorus to any substantial level. After eight weeks the two parallel systems developed a sludge containing 2.1% phosphorus. At this point the application of external electrical current was started in the electrified system.

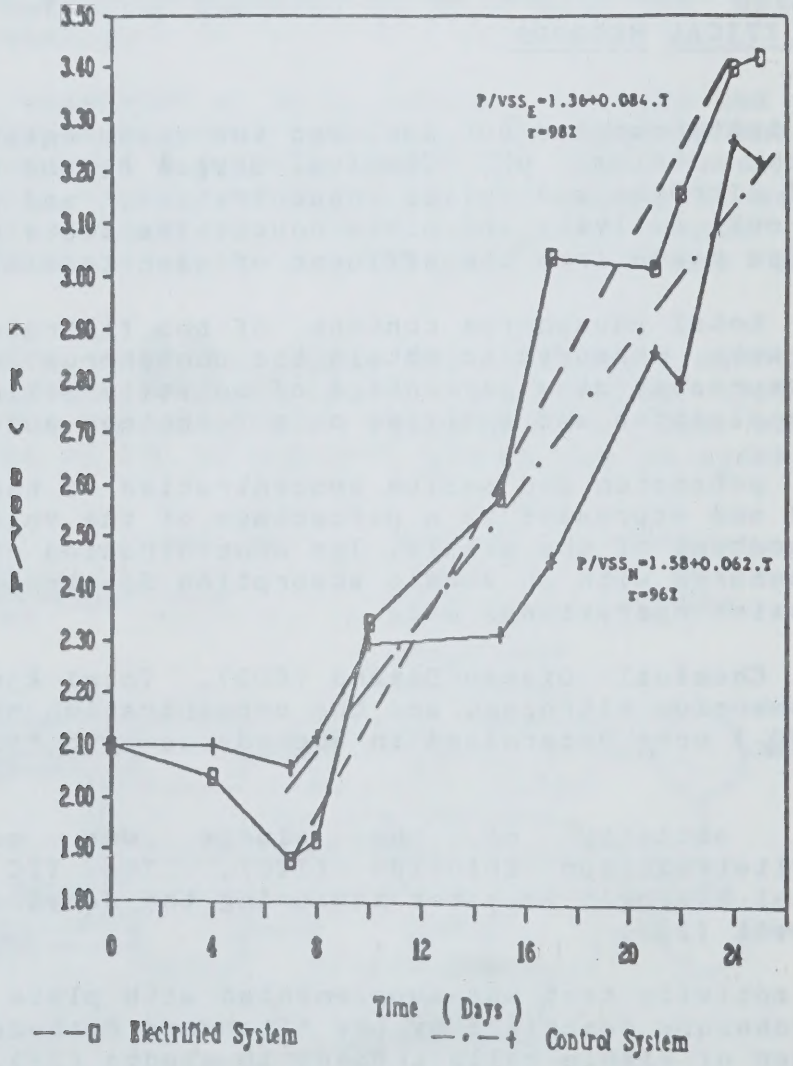

Fig. 2 Phosphorus Content of Sludge during Start Up

During the start-up period of the biological system the electric current increased the rate of phosphorus accumulation in sludge by 30% from 0.62 mg P/g VSS/d to 0.84 mg P/g VSS/d as it can be seen in Fig.2.

Under steady state conditions, as can be seen in Fig.3, the application of external electric current increased the phosphorus content of sludge after the aerobic step by 11% from 2.60% to 2.89%. The difference is statistically significant at 99% confidence level.

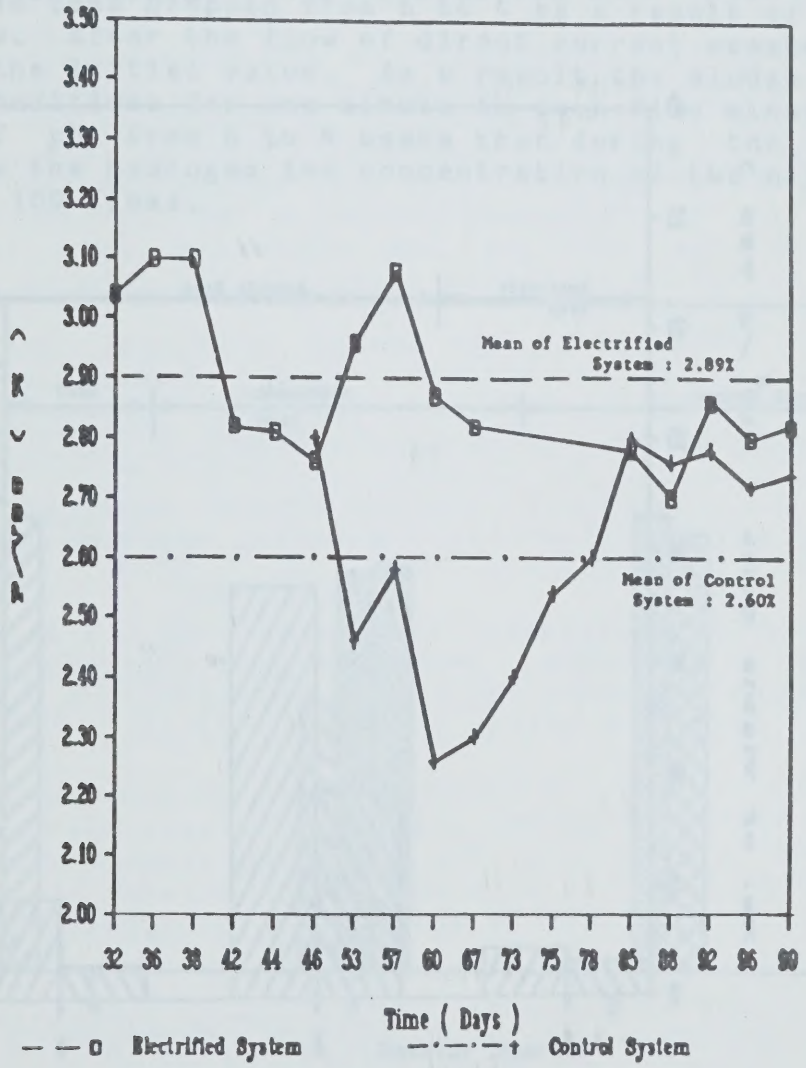

Fig. 3 Excess Phosphorus Uptake during Steady-State

It was observed that the increased uptake of phosphorus under aerobic conditions was not the result of increased release of phosphorus during the anaerobic step.

The increase observed in the amount of phosphorus stored in the sludge indicated that the direct current might alter some of the metabolic reactions or composition of the activated sludge.

As a result of the electrical stimulus the electrified system had 2 to 5 times less viable cells present in the sludge than did the control. The lower cell count obtained for electrified sludge might be the result of the direct current, killing off those bacteria which were not able to store excess energy. Refer to Fig.4.

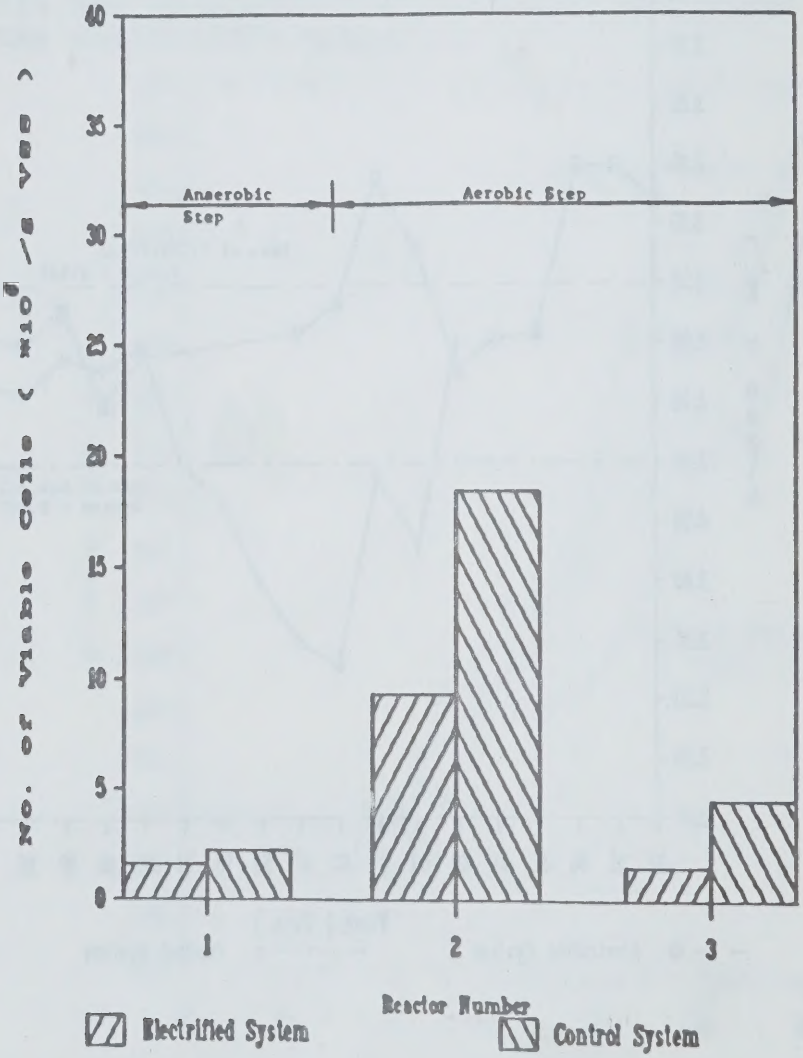

Fig. 4 Viable Cell Count

As a result of the reduced and possibly changed bacterial population the electrified sludge showed a higher specific cell activity than did the control. Similarly, there was a significant difference between the electrified and control systems in potassium ion uptake and release.

The reduction in viable cell count did not effect the COD removal efficiency nor the nitrification - denitrification processes of the biological system.

The electric current had, however, a direct effect on the pH values as it can be seen in Fig.5. The pH of the mixed liquor in the anaerobic tank dropped from 6 to 4 as a result of the direct current flow. After the flow of direct current ceased, the pH returned to the initial value. As a result the sludge was under pH stress conditions for one minute in each five minutes cycle. The drop of pH from 6 to 4 means that during the period of current flow the hydrogen ion concentration of the mixed liquor increased by 100 times.

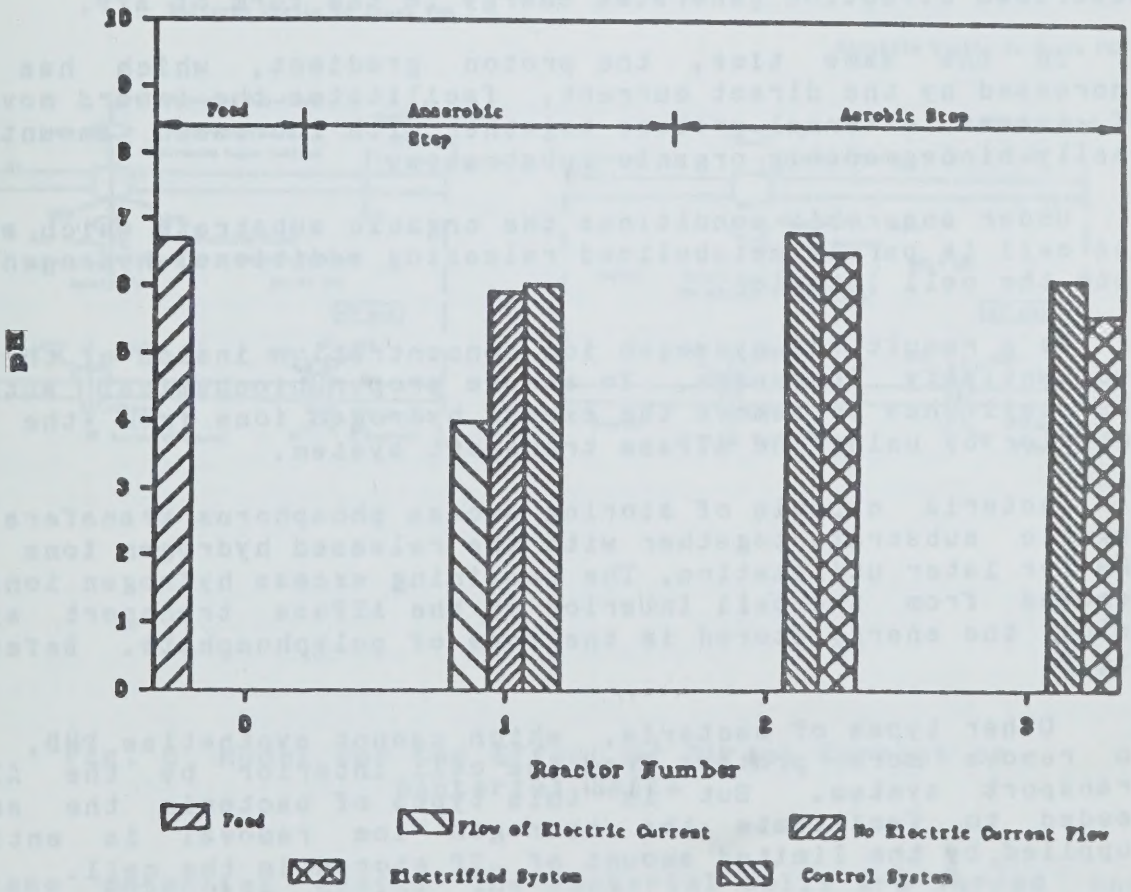

Fig.5 pH of Mixed Liquor

MODEL FOR THE EFFECT OF DIRECT CURRENT ON BIOLOGICAL PHOSPHORUS UPTAKE

The application of direct current increased the phosphorus content of the sludge, changed tha pattern of potassium ion movements, lowered the pH values during the flow of electric current, and reduced the number of viable cells present in the sludge.

From the results it seems, that the drop in pH values during electrification might have initiated the other changes observed.

The drop in pH increased the proton concentration in the supernatant. The high proton concentration outside the bacterial cells changed the electrochemical potential of the cell membrane which can be restored by discharging some of the intracellular potassium ions. The potassium ions can be removed from the cell interior via the hydrogen-potassium antiporter system. The movement of the hydrogen and potassium ions in the above described direction generates energy in the form of ATP.

In the same time, the proton gradient, which has been increased by the direct current, facilitates the inward movement of excess external protons together with increased amounts of easily biodegradable organic substrates.

Under anaerobic conditions the organic substrate which enters the cell is partly metabolized releasing additional hydrogen ions into the cell interior.

As a result the hydrogen ion concentration inside of the cell substantially increases. To ensure proper biochemical activity the cell has to remove the excess hydrogen ions from the cell interior by using the ATPase transport system.

Bacteria capable of storing excess phosphorus transfers the organic substrate together with the released hydrogen ions into PHB for later utilization. The remaining excess hydrogen ions are removed from the cell interior by the ATPase transport system using the energy stored in the form of polyphosphate. Refer to Fig.6.

Other types of bacteria, which cannot synthetise PHB, have to remove more protons from the cell interior by the ATPase transport system. But in this types of bacteria the energy needed to facilitate the hydrogen ion removal is entirely supplied by the limited amount of ATP stored in the cell.

As the electric field is removed the hydrogen ion concentration of the mixed liquor drops, thus changing the

During Electrification

After Electrification

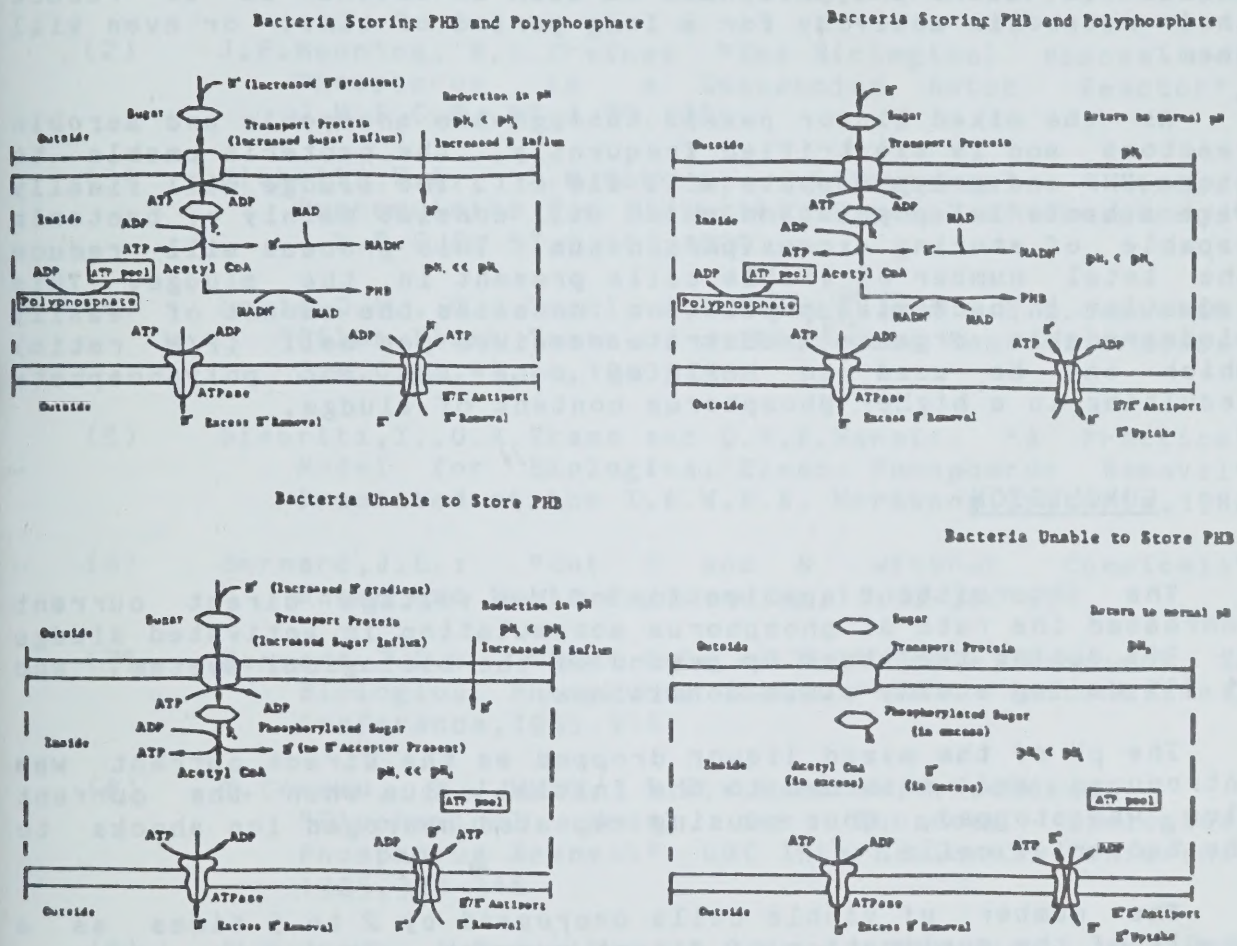

Fig. 6 Model for the Effect of Direct Current on Bacterial Cells

membrane potential again. The bacterial cells are forced once again to restore the normal membrane potential, by taking up some of the previously discharged potassium ions. The uptake of potassium ions and the release of protons require energy, which

further reduces the energy available for the bacterial cells.

By repeating this process several times, as occured during this study, the stored energy level of the cells will be reduced.

This process can exhaust the supply of energy from cells unable to store polyphosphate to such an extent as to reduce their metabolic activity for a long period of time, or even kill them.

As the mixed liquor passes through the anaerobic and aerobic reactors and is electrified frequently, the bacteria unable to store PHB and polyphosphate will die off. The sludge will finally have a bacterial population which will consist mainly of bacteria capable of storing excess phosphorus. This process will reduce the total number of viable cells present in the sludge. This reduction in bacterial population increases the amount of easily biodegradable organic substrate received per cell (F/M ratio) which can be used to build up more PHB and polyphosphate resulting in a higher phosphorus content of sludge.

CONCLUSION

The intermittent application of low voltage direct current increased the rate of phosphorus accumulation in activated sludge by 30% during the start up period of the biological system, and by 11% during steady state conditions.

The pH of the mixed liquor dropped as the direct current was introduced and returned to the initial value when the current flow was stopped, thus causing repeated hydrogen ion shocks to the bacterial cells.

The number of viable cells decreased by 2 to 5 times as a result of the application of direct current. It may be, however, that the reduction in viable cell count was the immediate result of pH shock.

As a result of the reduction of the number of viable bacterial cells present, the food to microorganism ratio increased. More food means that more PHB can be built up which makes it possible to accumulate more polyphosphate resulting in enhanced phosphorus uptake. Thus, the application of intermittent low voltage direct current increased the F/M ratio to such a value which was favorable for excess phosphorus uptake.

This treatment technique may be useful at treatment plants which receive little organic load, diluted wastewater or wastewater with high nitrogen to COD ratio.

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ROTATING BIOLOGICAL CONTACTOR TECHNOLOGY

By K.L. Murphy[1]

1.0 INTRODUCTION

The rotating biological contactor (RBC) may be defined as a biological process in which a media for biological attachment is rotated alternatively between the wastewater and the atmosphere above the wastewater. Microorganisms attached to the disk obtain substrate and nutrients from their exposure to the wastewater and oxygen from their exposure to the atmosphere. The turbulence associated with the movement of the disk provides the requisite mixing to ensure adequate contact between attached microorganisms and the substrate and nutrients in the wastewater. Generally, a series of compartments or stages of disks is provided to obtain the required degree of treatment. Recently, RBC units have been operated completely submerged or in an atmosphere devoid of oxygen to provide an anoxic or anaerobic environment for the attached microorganisms to facilitate nitrogen removal (denitrification) or methane production. Experience with these anoxic or anaerobic units has been obtained principally from bench or pilot scale studies and will not be included in this review which concentrates on large pilot or full scale results.

2.0 HISTORICAL DEVELOPMENT

Reports on the development of partially submerged rotating disks appeared in the literature as early as 1929 (BEAK, 1973). Development of a commercial product was undertaken in Germany in the 1950's and was marketed by J.C. Stengelin, Tuttlingen, West Germany. This company initiated licensing agreements in a number of European countries as well as with Allis Chalmers Corp. for the U.S. and Canadian markets. In 1970 the business was sold to Autotrol Corp.

3.0 INTRODUCTION TO NORTH AMERICA

The first RBC units were constructed with flat 0.5 in thick, 6.5 to 10.0 ft diameter expanded polystyrene sheets. Marketing efforts were not particularly successful until the subsequent development of a corrugated high density polyethylene media by Autotrol in 1972. The major advantage of polyethylene is the ability to form it into various configurations with a thickness of only 0.04 to 0.06 inches. This has enabled 100,000 to 180,000 ft^2 of surface area an increase of 70 to 150% in disk surface area, on a single 27ft shaft and has allowed the RBC process to become more cost competitive with other biological treatment systems. Table 3.1 lists the manufacturers who have installed the majority of the installations in North America.

The principal differences among the different vendors' products are associated with media configuration, fabrication, attachment, and support shaft design. Differences also exist in drive systems and type of bearing support.

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Various media configurations or corrugation patterns have been selected by the manufacturers, each with its own claimed advantages. One manufacturer, CMS Rotodisk, uses flat disks of 3 mesh, medium density, polyethylene. All other manufacturers utilize some form of a corrugated disk with the exception of Walker Process. Walker use corrugated polyethylene in a continuous spiral wrapping to produce a monolithic media eliminating the requirement for media supports.

Corrugations used by most of the manufacturers add stiffness to the sheets and enable these sheets to be formed with diameters as large as 12 ft. This increases the available surface area by 15 to 20 percent. Corrugations cause the wastewater to follow a tortuous path through the media, thus increasing wastewater exposure time to the air for greater oxygen transfer in the atmospheric sector of the rotational cycle. Finally, the corrugations are used as spacers to keep the sheets separated.

Standard density media are normally used in the leading stages of an RBC train. Standard density media are defined as media with a surface area of 100,000 sq ft supported on or from a 27-ft shaft in which the media diameter is approximately 12 ft. By reducing the space required for the repeating plastic corrugations by 33 percent, the available surface area can be effectively increased by 50 percent; this results in shafts with 150,000 sq ft of media, commonly called high density media. Some manufacturers are now also offering media with densities of 120,000 and 180,000 sq ft per27-ft long shaft for increased design flexibility.

RBC manufacturers employ various methods for supporting plastic media from their shafts. Clow, Crane-Cochrane, and Lyco rely on a coated steel or stainless steel radial arm system to support the plastic media. Envirex employs plastic media hubs that fit onto its square and octagonal shafts with the plastic media sheets thermally welded to the hubs. Walker Process attaches the plastic media to the shaft with an epoxy bonding agent and stainless steel strips with the media then spirally wound onto the shaft in 35-in. wide strips.

RBC shafts are used to support and rotate the plastic media. Maximum shaft length is presently limited to approximately 27 ft, with 25 ft occupied by media. Shorter shaft lengths are also available. The shafts are fabricated from steel and are covered with a protective coating suitable for water and high humidity service. Proper protective coating procedure requires sand blasting of the steel prior to the coating application. A coal tar epoxy is normally used as the protective coating with a minimum film thickness of 14 mils (0.014 in.). Each manufacturer designs its own shape, size, and thickness of shaft (Table 3.1). The wall thickness of the shaft is governed by structural requirements, and the shape is highly dependent on the method the manufacturer employs in supporting the plastic media from the shaft. All manufacturers utilize a shaft that differs from the others in either thickness, size, or shape, or in some cases all three. Structurally, these differences are readily apparent as shown in Figure 3.1.

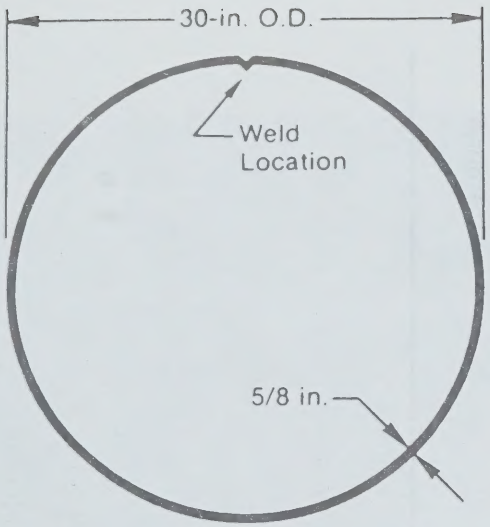
RBC manufacturers specify factory-assembled drive packages for all mechanical drive equipment, consisting of motors, speed reducers, and drive systems. Reduction of motor output speed down to approximately 1.6 rpm can be accomplished through the output use of various combinations of multi-V-belts, gear boxes, and chain-and-sprocket units.

Envirex offers an air driven RBC unit. The air drive assembly consists of 4- or 6-in. deep plastic cups welded around the outer perimeter of the media and an air header

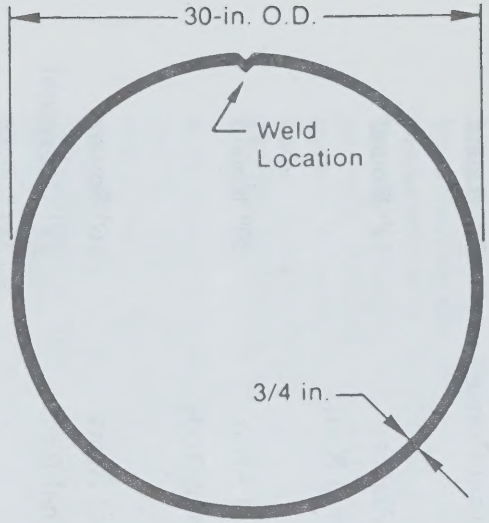
TABLE 3.1 CHARACTERISTICS OF COMMERICAL RBC's

Manufactuer	Media	Media Support	Shaft
Autotrol/Envirex	Disk Closed Form	Self Supporting	16" Square 30" Octagonal
Bioshaft	Disk	Radial Arms Peripheral Rods	
Clow Envirodisc	Disk Open Form	Radial Arms External Frame	30" Round
CMS Rotodisc	Disk Open Form	End Plates Internal Rods	14" Round*
Crane Cochrane	Disk	Radial Arms Internal Rods	30" Round
Lyco/Hormel	Disk Closed Form	Radial Arms Internal Rods	16" Square 24" Octagonal 28" Round
Walker	Spiral Wound	-	30" Round

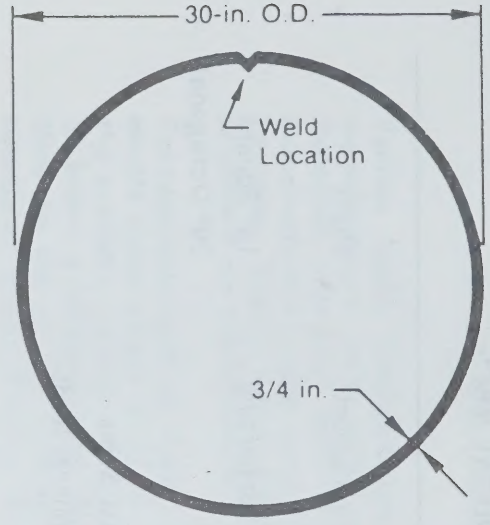
* Limited to 62,500 ft of surface area



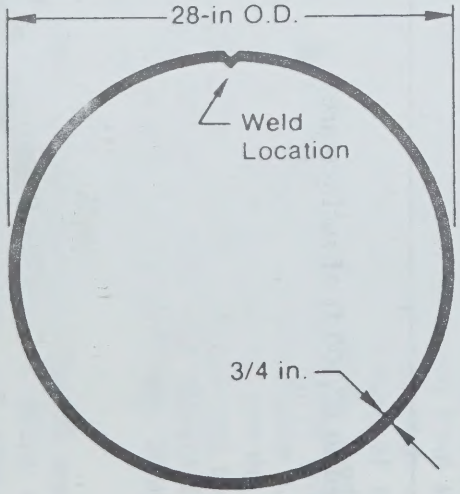
CLOW



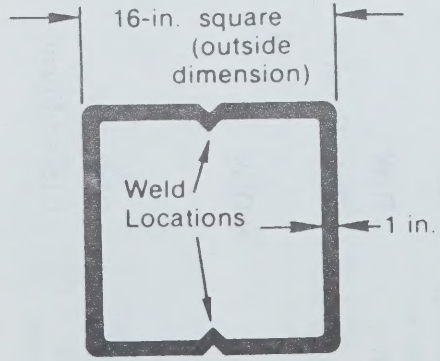
WALKER PROCESS



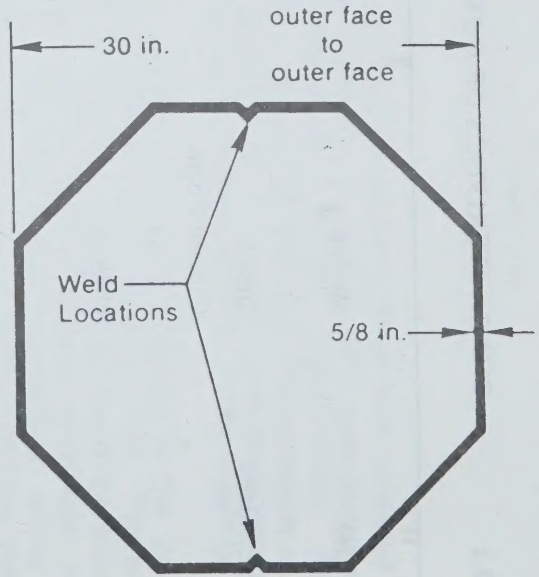
CRANE-COCHRANE



LYCO



ENVIREX/AUTOTROL



ENVIREX

FIGURE 3.1 Cross-sections of RBC Shafts

placed below the media. Air is released at a pressure of 3 to 4 psig into the attached cups, creating a buoyant force that causes the shaft to turn. Approximately 20 to 30 percent of the air is not captured by the cups and escapes into the radial passages where it flows upward through the media.

4.0 DESIGN CRITERIA

As the practical application of the RBC has evolved over the last 15 to 20 years in Europe and North America, several different methodologies for design have been proposed. Some researchers have attempted to adopt a traditional approach used with suspended-growth biological processes, by defining an organic loading or F:M ratio. A theoretical mixed liquor volatile suspended solid's concentration (MLVSS) is determined by scraping the biomass from a known area of media, calculating the mass of volatile solids per unit area, and multiplying this value by the total media surface available. This total mass of volatile solids is then divided by the liquid volume of the RBC to arrive at the MLVSS. This approach has not enjoyed wide acceptance for RBC design, perhaps because of the difficulty in measuring or predicting the MLVSS concentrations necessary for the calculations.

Empirical models have been widely used to describe or predict RBC performance. Joost (1969) proposed the following relationship based on substrate concentration, temperature and residence time:

$$\frac{\%\ BOD_5\ \text{removed}}{\text{Stage}} = K \times C^a \times R^b \times T^c \times S^d$$

where: K is a specific parameter for each waste.
C is the substrate concentration,
R is the reactor constant,
T is the temperature,
S is the hydraulic residence time, and
a,b,c,d are dimensionless parameters.

This model suffers from two handicaps typical of many of those developed: first because it was derived using bench scale and small pilot scale data, and therefore its results may not apply to the larger, full scale RBC's; and second, and more importantly, because for every RBC a separate set of parameter values must be derived. Joost did not determine whether a common set of parameters, generally applicable to municipal wastewater treatment, could be developed.

Mechanistic models have been developed that consider substrate transport and metabolism and oxygen transport and depletion within the biofilm as a function of liquid film thickness, biofilm thickness, rotational speed, mixing with the bulk liquid, etc. The relative rates of both transport and diffusion must be considered in any generalized attempt to construct deterministic mathematical models that predict accurately RBC performance. Discussion of the relative influence of these rates is provided by Grady and Lim (1980). When all of the factors affecting substrate removal and oxygen transfer are considered, it is apparent that there is no a priori reason to expect substrate removal from the bulk liquid in an RBC system to follow any simple mathematical model.

The design approach recommended by RBC manufacturers utilizes average design flows and BOD_5 concentrations to determine media requirements. This implicitly

relates organic loading to RBC effluent quality. This has been supported by the results of a pilot scale study (0.5 m diam) conducted by Poon and Mikucki (1979), who used a factorial design to determine which variables among flow, BOD_5 concentration, recirculation, chloride concentration and organic loading rate per unit area, had the greatest effect on BOD_5 removal efficiency. The organic load was found to be the variable that most closely correlated with effluent BOD_5 concentration. None of these variables had a large effect on the percentage of BOD_5 removal.

Further support comes from studies conducted at the Wastewater Technology Centre Burlington (Murphy and Wilson 1980). A 2.0 m RBC was tested using a full factorial design on the variables flow. Total organic carbon (TOC) concentration and loading, and total kjehdahl nitrogen (TKN) concentration and loading. TOC load had the greatest impact on effluent TOC concentration when operated in a carbon removal mode. As a close correlation between TOC and BOD_5 was obtained, BOD_5 loading also would be the major design parameter in this operational mode. TKN loading had the major impact on the degree of nitrification attained where nitrification was the objective.

RBC manufacturers generally recommend a minimum of 4 stages where low effluent BOD_5 values or relatively high BOD_5 removals are required. Similarly, nitrification can be accomplished most effectively after BOD_5 reduction is accomplished, generally after the second or third stage. Use of different tank configurations and tank volume to media surface area ratios will cause differences in mixing conditions, amounts of dead space, hydraulic retention time, and oxygen transfer rates into the bulk liquid. Although these could affect comparative results from different makes of RBC, no supporting quantitative data are available .

Murphy and Wilson (1980) demonstrated that under identical operating conditions a 0.5 m RBC averaged 16 percent higher mass removals of COD per unit area and time than did a 2.0 m unit. The difference in removals was statistically significant at a confidence level of 95 percent. Under combined BOD_5 removal plus nitrifying conditions, mass removals of filterable TKN were similar for both units. This would indicate that the oxygen transfer capability of the 2.0 m unit may be limiting at high loadings. When using pilot scale data for design of a full scale RBC, a scale-up factor would be required. Similar results on nitrification were reported (Brenner et al, 1984). They cited pilot scale RBC nitrification rates in the literature 1.5 to 2.5 times peak full scale rates and explained the difference by the limitation in mass transfer associated with the reduced rotational velocities used by the full scale units as compared to pilot scale units.

5.0 EARLY PROBLEMS

A number of operational problems were associated with early full scale installations despite the simplicity of operation associated with RBC treatment. These could be identified as equipment failure caused by either excessive growth of attached microorganisms or eccentric loading of microorganisms on the disk surface or a poorer than predicted effluent quality.

Equipment Reliability

An EPA sponsored study (Brenner et al, 1984) of 17 early installations indicated that for the 153 units installed there were 41 broken shafts, 7 media attachment

failures, and 16 bearing failures. Only three installations had not suffered any form of failure.

The most serious equipment failure is a structural break in the RBC shaft. This failure usually results in damaged media and requires the removal of the entire unit and subsequent replacement with a new shaft along with salvaged and/or new media. Depending on the site layout this may require removal of protective media covers or some reconstruction of the process building. Most early shaft failures were caused by unsatisfactory welding practice or fatigue failure from the applied cyclic loading.

Six of the plants reported media failure, shifting media, brittleness and breakage. Degradation of the polyethylene media can be caused by exposure to heat (79°F) concentrated solvents, or UV degradation. Media breakage may result from poor hub design and media welding procedures or inadequately designed radial arm systems. Replacement of media may or may not require removal of the shaft depending on the manufacturers design.

Only 3 plants reported failure of drive components.

Effluent Quality

As indicated in Section 4.0 the design approach for RBC used in North America has related effluent concentration to the average influent loading. Despite the use of conventional loadings some RBC installations did not yield the anticipated effluent quality. This reduction in effluent quality has been shown to result from inadequate pretreatment and variations in influent loading.

Pretreatment

The US EPA as a result of their study of RBC facilities (Brenner et al,1984) state raw municipal wastewater should not be applied to an RBC system. Grit removal and either primary clarifiers or screens are necessary to remove materials that may settle in the RBC tankage or plug the RBC media. High influent grease loads require the use of primary clarifiers or an alternate system for grease removal. Effective preliminary treatment is essential for good operation.

In the study undertaken at WTC Burlington (Murphy and Wilson, 1980) a direct comparison was undertaken on the effect of primary clarification on RBC effluent quality. Under similar BOD_5 and hydraulic loadings primary clarification reduced the average suspended solid concentration in settled RBC effluent by approximately 50% (Figure 5-1). As the average BOD_5 associated with the effluent suspended solids averaged 0.3 g BOD/g TSS (Figure 5.2) primary clarification translates into a direct reduction in total effluent BOD_5.

Flow Equalization

The literature reports that RBC's can withstand both hydraulic and organic shock loads avoiding washouts and other upsets typical of activated sludge systems. The results of recent studies (Murphy and Wilson, 1980) conducted at WTC on the effects of diurnal variations support this conclusion. However, it is not true that the RBC can absorb these shock loads with little or no effect on treatment efficiency. The hourly data from both the BOD_5 removal and the combined BOD_5 removal plus nitrification study in Figures 5.3 and 5.4 demonstrate that shock loads

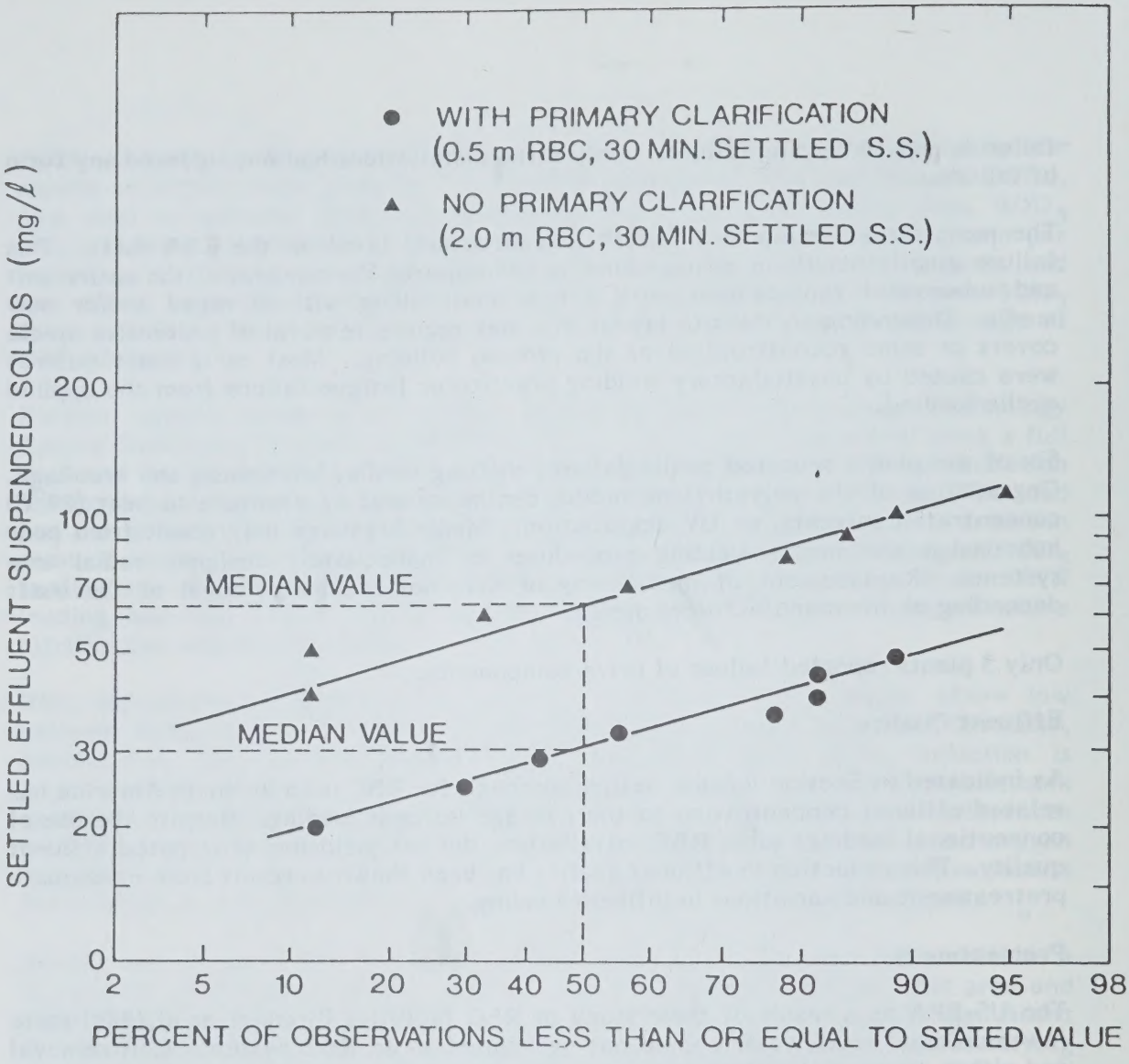


FIGURE 5.1 Effluent Suspended Solids Concentration With and Without Primary Clarification (After Murphy and Wilson, 1980)

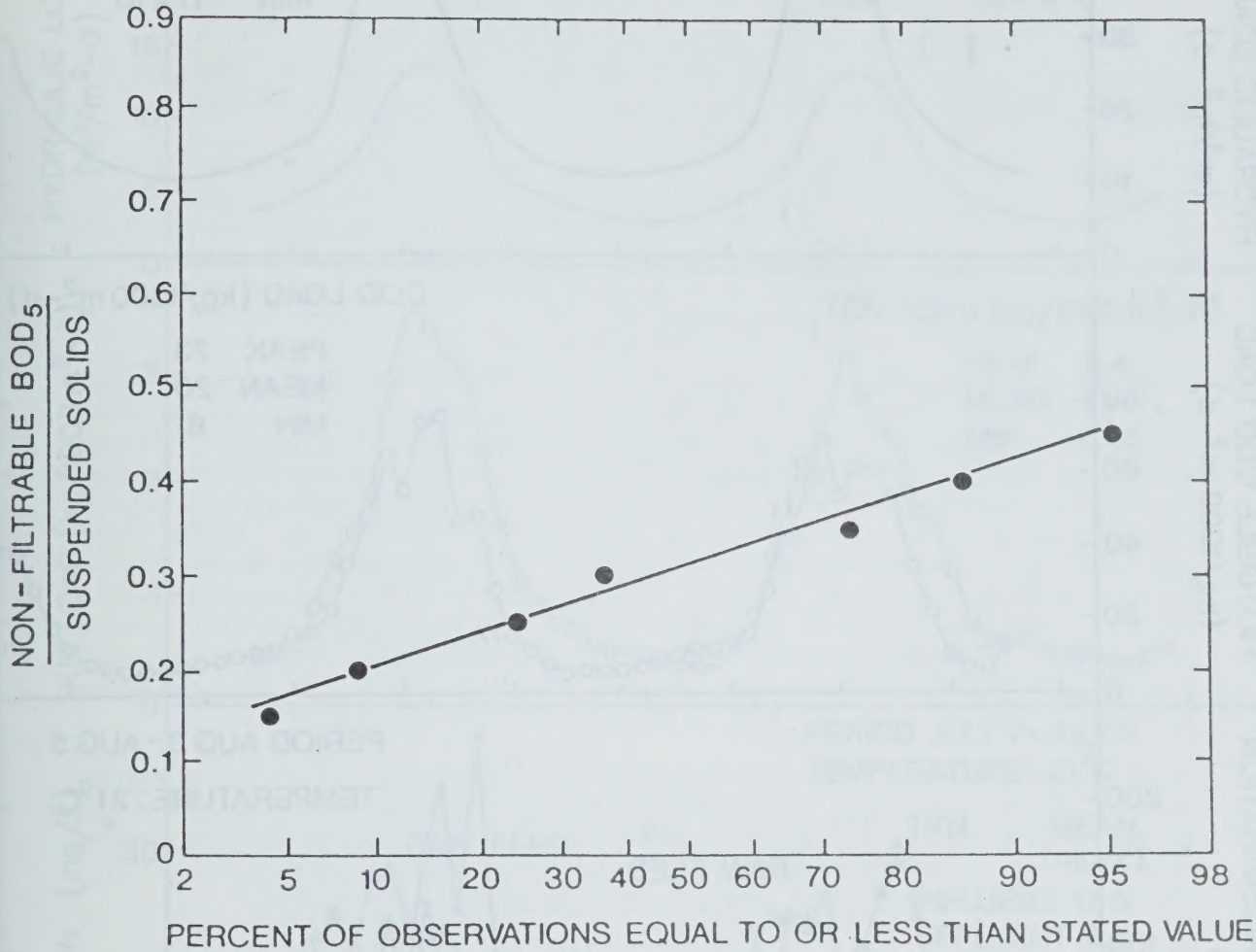


FIGURE 5.2 Ratio of Non-filterable BOD_5 to Suspended Solids in Settled Effluent (2.0 m RBC) (After Murphy and Wilson, 1980)

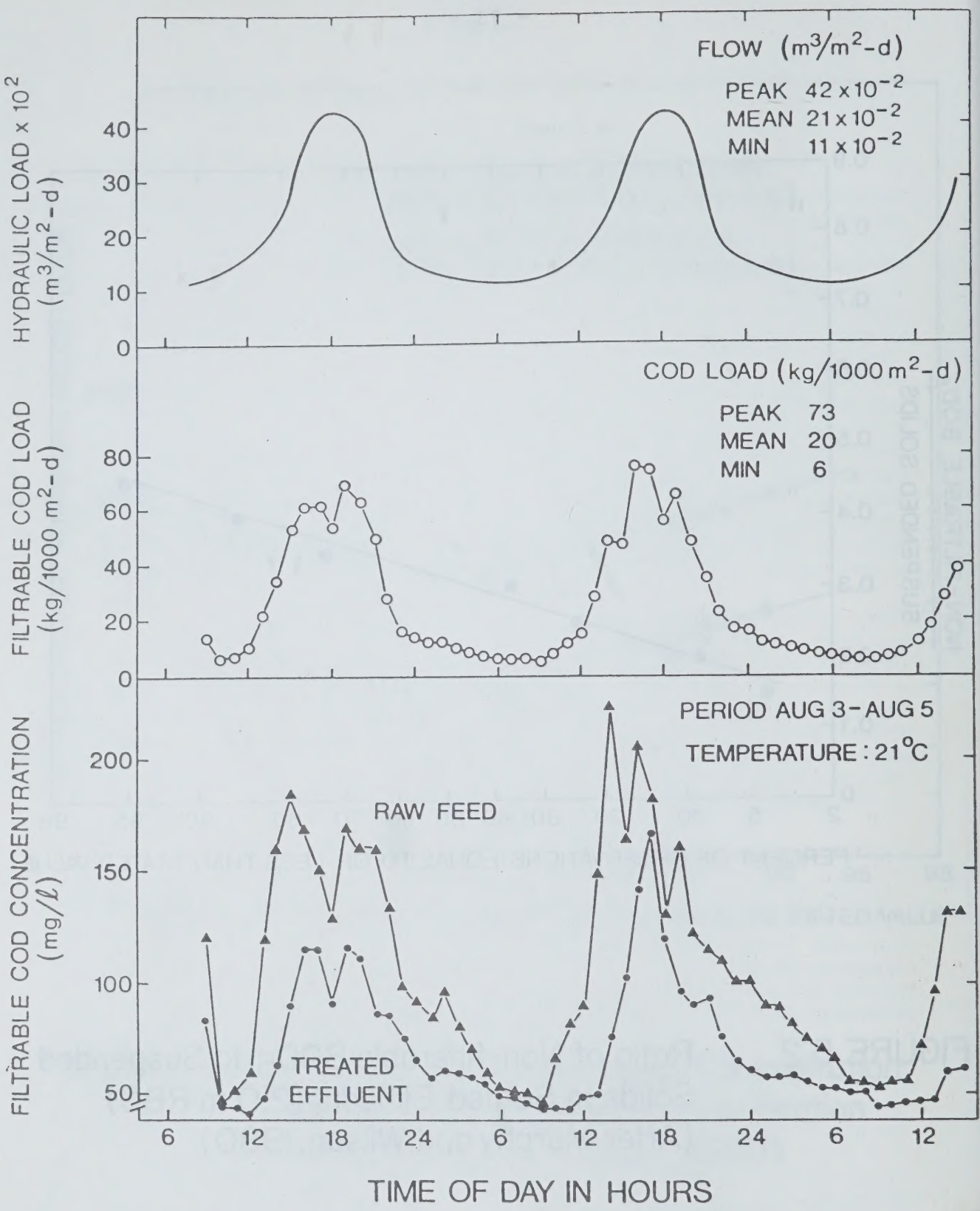


FIGURE 5.3 Effect of Diurnal Flow and Concentration Variation on RBC Performance (0.5 m RBC in the Carbon Removal Mode) (After Murphy and Wilson, 1980)

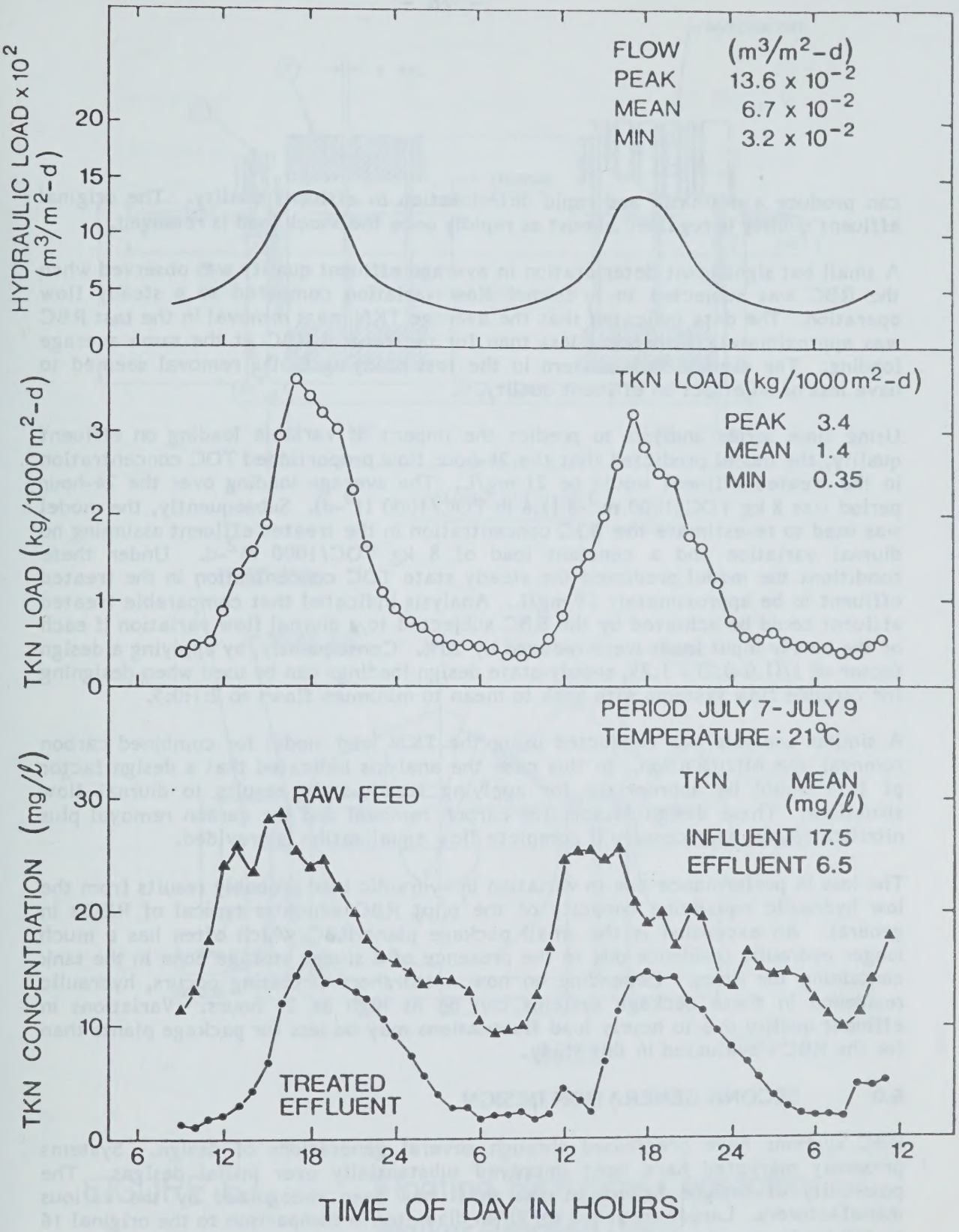


FIGURE 5.4 Effect of Diurnal Flow and Concentration Variation on RBC Performance (0.5 m RBC in the Carbon Removal plus Nitrification Mode)

can produce a dramatic and rapid deterioration in effluent quality. The original effluent quality is regained almost as rapidly once the shock load is removed.

A small but significant deterioration in average effluent quality was observed when the RBC was subjected to a diurnal flow variation compared to a steady flow operation. The data indicated that the average TKN mass removal in the test RBC was approximately 20 percent less than for the control RBC at the same average loading. The diurnal flow pattern in the test studying BOD_5 removal seemed to have less of an effect on effluent quality.

Using time series analysis to predict the impact of variable loading on effluent quality, the model predicted that the 24-hour flow proportioned TOC concentration in the treated effluent would be 21 mg/L. The average loading over the 24-hour period was 8 kg TOC/1000 m^2-d (1.6 lb TOC/1000 ft^2-d). Subsequently, the model was used to re-estimate the TOC concentration in the treated effluent assuming no diurnal variation and a constant load of 8 kg TOC/1000 m^2-d. Under these conditions the model predicted the steady state TOC concentration in the treated effluent to be approximately 19 mg/L. Analysis indicated that comparable treated effluent could be achieved by the RBC subjected to a diurnal flow variation if each of the hourly input loads were reduced by 20%. Consequently, by applying a design factor of 1/(1.0-0.2) = 1.25, steady-state design loadings can be used when designing for variable flow systems with peak to mean to minimum flows to 2:1:0:5.

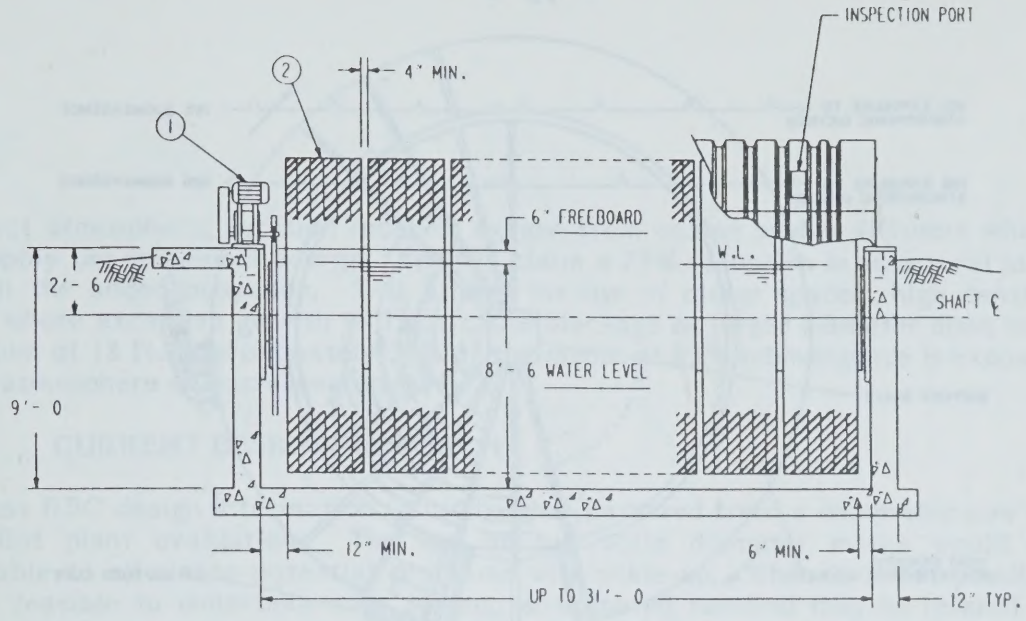
A similar analysis was conducted using the TKN load model for combined carbon removal plus nitrification. In this case the analysis indicated that a design factor of 1.35 would be appropriate for applying steady-state results to diurnal flow situations. These design factors for carbon removal and for carbon removal plus nitrification are unnecessary if complete flow equalization is provided.

The loss in performance due to variation in hydraulic load probably results from the low hydraulic equalizing capacity of the pilot RBC which is typical of RBC's in general. An exception is the small package plant RBC which often has a much longer hydraulic residence due to the presence of a sludge storage zone in the tank containing the discs. Depending on how much short circuiting occurs, hydraulic residence in these package systems can be as high as 24 hours. Variations in effluent quality due to hourly load fluctuations may be less for package plants than for the RBC's evaluated in this study.

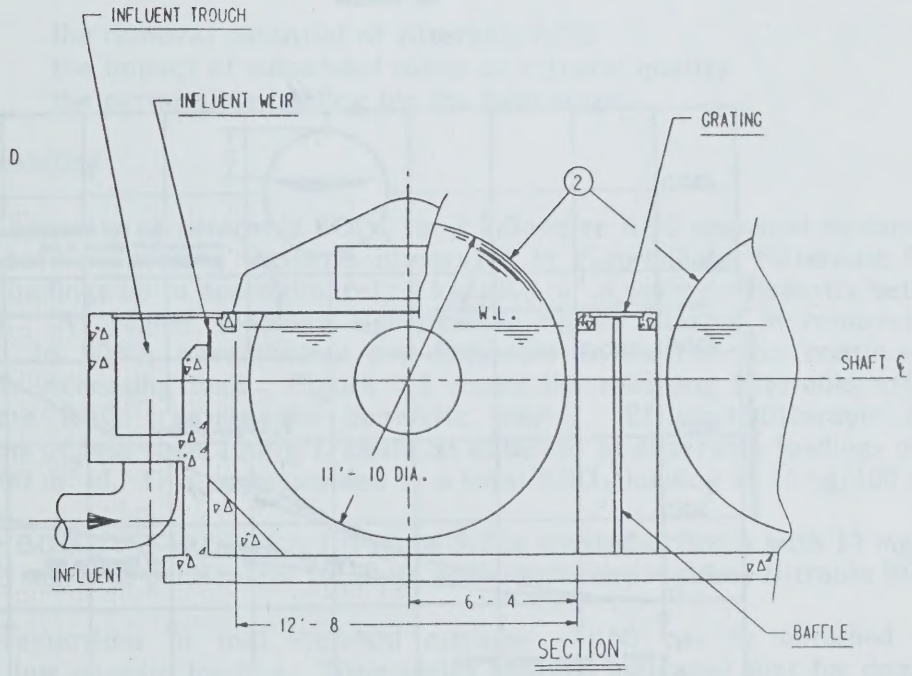
6.0 SECOND GENERATION DESIGN

RBC systems have progressed through several generations of design. Systems presently marketed have been improved substantially over initial designs. The possibility of fatigue failure of the shaft has been recognized by the various manufacturers. Larger shafts 28 to 30" in diameter in comparison to the original 16 to 24" shafts are the recognized norm. Standards of the American Welding Society AWS are used for design. Not withstanding design improvements, load cells should be located at least on the initial shafts to monitor biomass accumulation and potential shaft overloading. Walker Process is now marketing a Floating Biological Contactor (Figure 6.1). The FBC has an even larger central shaft (48 in diameter) which is completely submerged so that the buoyant effect completely supports the load of the shaft, media and developed biomass.

Envirex also markets a submerged RBC (Figure 6.2) where 70 to 90% of the media is submerged in contrast to the 40% in a conventional RBC. The media not exposed

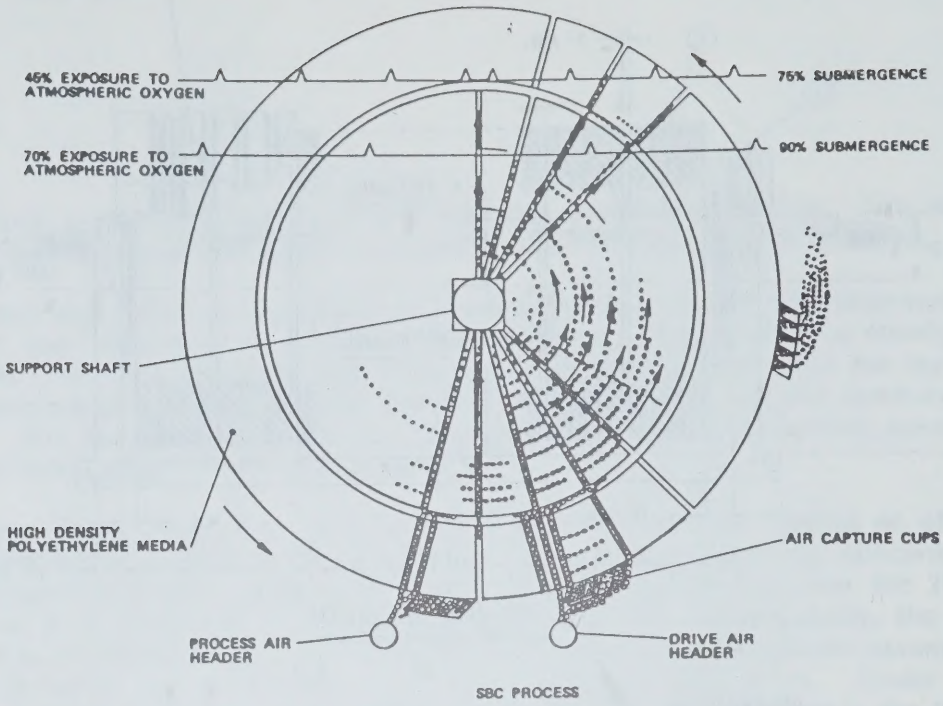


SECTIONAL ELEVATION



SECTION

FIGURE 6.1 Floating Biological Contactor-Walker Process Corp.



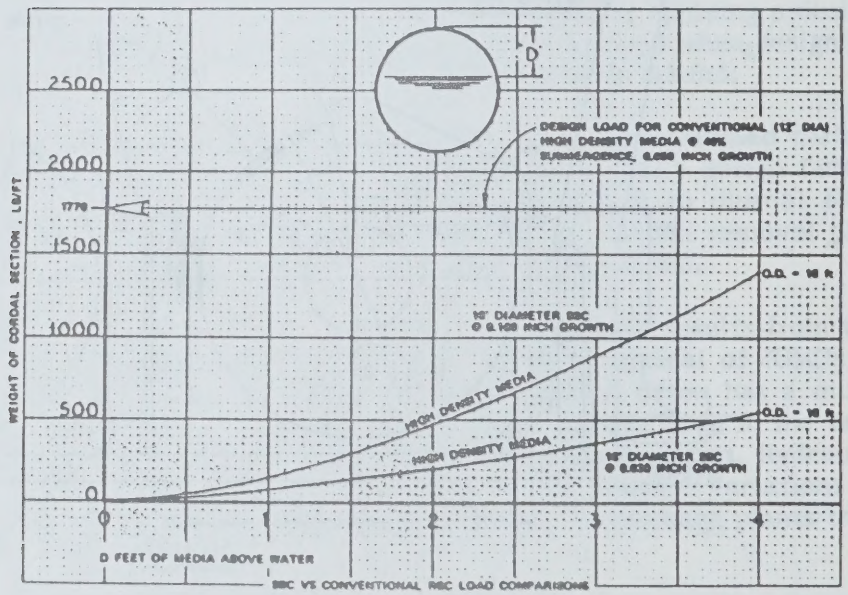


FIGURE 6.2 Submerged Biological Contactor-Envirex Inc.

to direct atmospheric aeration receives oxygen from coarse bubble diffusers which also supply the rotational energy. Envirex claim a 75% reduction in structural load through the added buoyancy. This allows the use of closer spaced, high density media where excessive growth will not cause blockage or larger diameter disks to a maximum of 18 ft. Approximately 55% of the media at 85% submergence is exposed to the atmosphere on each revolution.

7.0 CURRENT DESIGN APPROACH

The best RBC design information would be that gathered from a comprehensive on site pilot plant evaluation. The use of full scale diameter media would be preferable to eliminate potential problems with scale-up. When it is not possible and/or feasible to undertake such testing, anticipated removal may be related to mass loading. Because of the short hydraulic residence time generally found in RBC's primary treatment should be employed to minimize suspended solids loadings (Section 3). Consequently, RBC design should be based on

- o the removal potential of filterable BOD
- o the impact of suspended solids on effluent quality
- o the permissible loading for the first stage.

Removal Capability

The removal capacity of filterable BOD_5 for a 2.0 metre RBC operated on domestic sewage and food processing waste is illustrated in Figure 7.1. Filterable BOD_5 removals at loadings up to approximately 4 kg/1000 m^2.d were consistently between 60 and 80%. At higher loadings significantly higher scatter in removal was observed (50 to 80%); nevertheless the filterable BOD_5 removal continued to increase with increasing load. Figure 7.2 shows the resulting filterable effluent BOD_5 for the RBC treating the domestic waste. Effluent filterable BOD_5 concentrations of less then 10 mg/L could be obtained at filterable loadings of less then 3 kg/1000 m^2-d. This corresponded to a total BOD_5 loading of 10 kg/100 m^2 d.

Based on the BOD_5/TSS ratio of 0.3 (Figure 5.2) a treated effluent with 15 mg/L of TSS contain 5 mg/L of particulate BOD_5 in addition to any residual filtrable BOD_5.

Substantial reductions in toal kjehdahl nitrogen (TKN) can be obtained when operating at low organic loading. Time series analysis indicated that for domestic wastes TKN removal was dependent on TKN mass loading to the RBC. Murphy and Wilson (1980) found that TKN removal was consistently greater than 80% when loadings were less than 1.2 kg TKN/1000 m^2-d. At higher loadings, removal efficiency showed considerable scatter. Nitrite plus nitrate formation did not match TKN removal with effluent concentrations being consistently less then 50% of influent values, suggesting appreciable denitrification within the biomass.

First Stage Loading

A major constraint in the design of any RBC system is limiting the organic loading to the first stage(s) to values compatible with the oxygen transfer capability of the system. With excessive organic loadings, the biofilm thickness increases markedly and a white/grey biomass is frequently observed.

Organic overloading is, therefore, characterized by the growth of an excessively thick biological film on the media surface and can result in the following process problems:

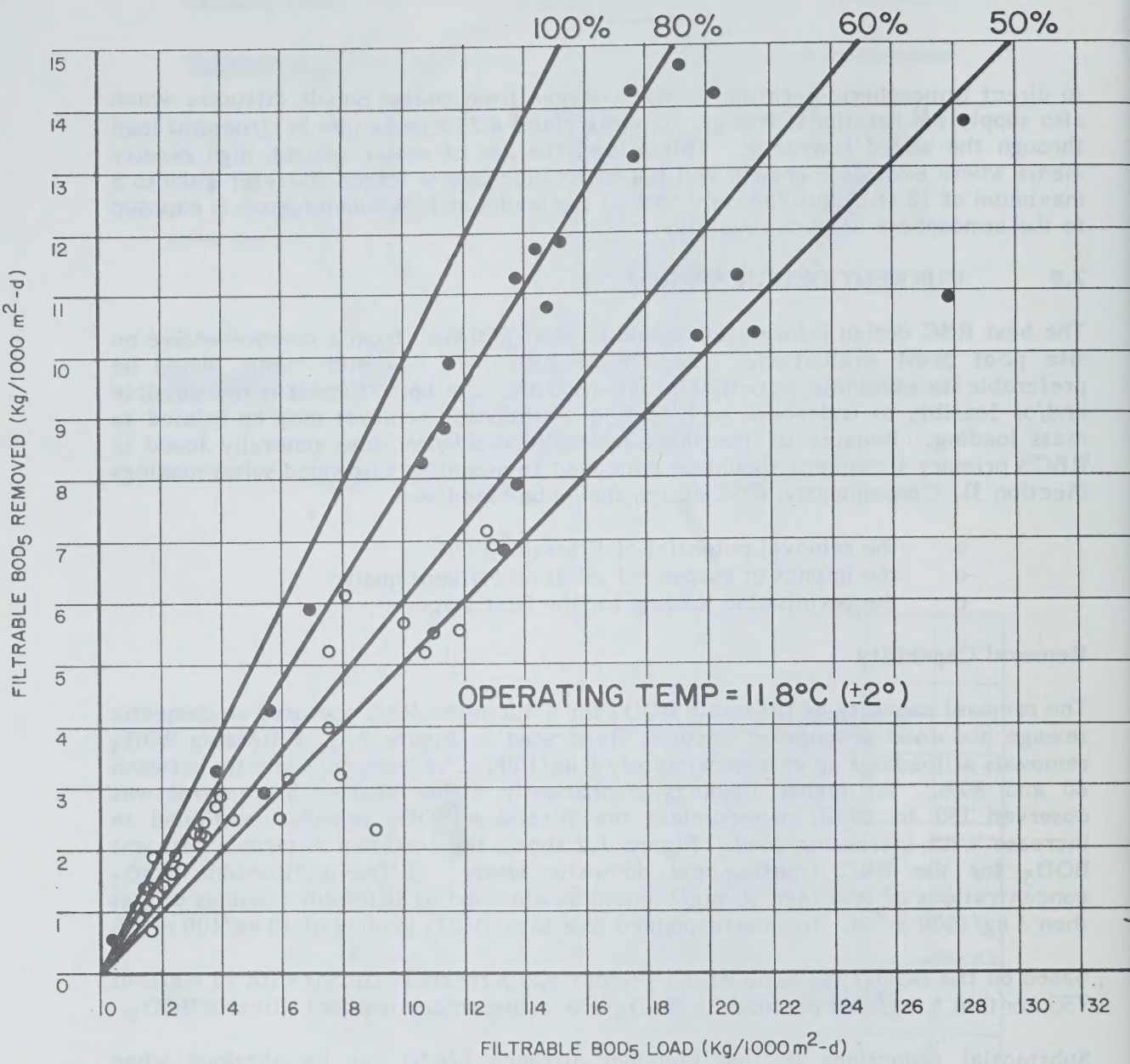


FIGURE 7.1 Filtrable BOD_5 Load vs. Filtrable BOD_5 Removed (2.0 m RBC)

NO PRIMARY CLARIFICATION
○ MUNICIPAL WASTE
● TOMATO WASTE

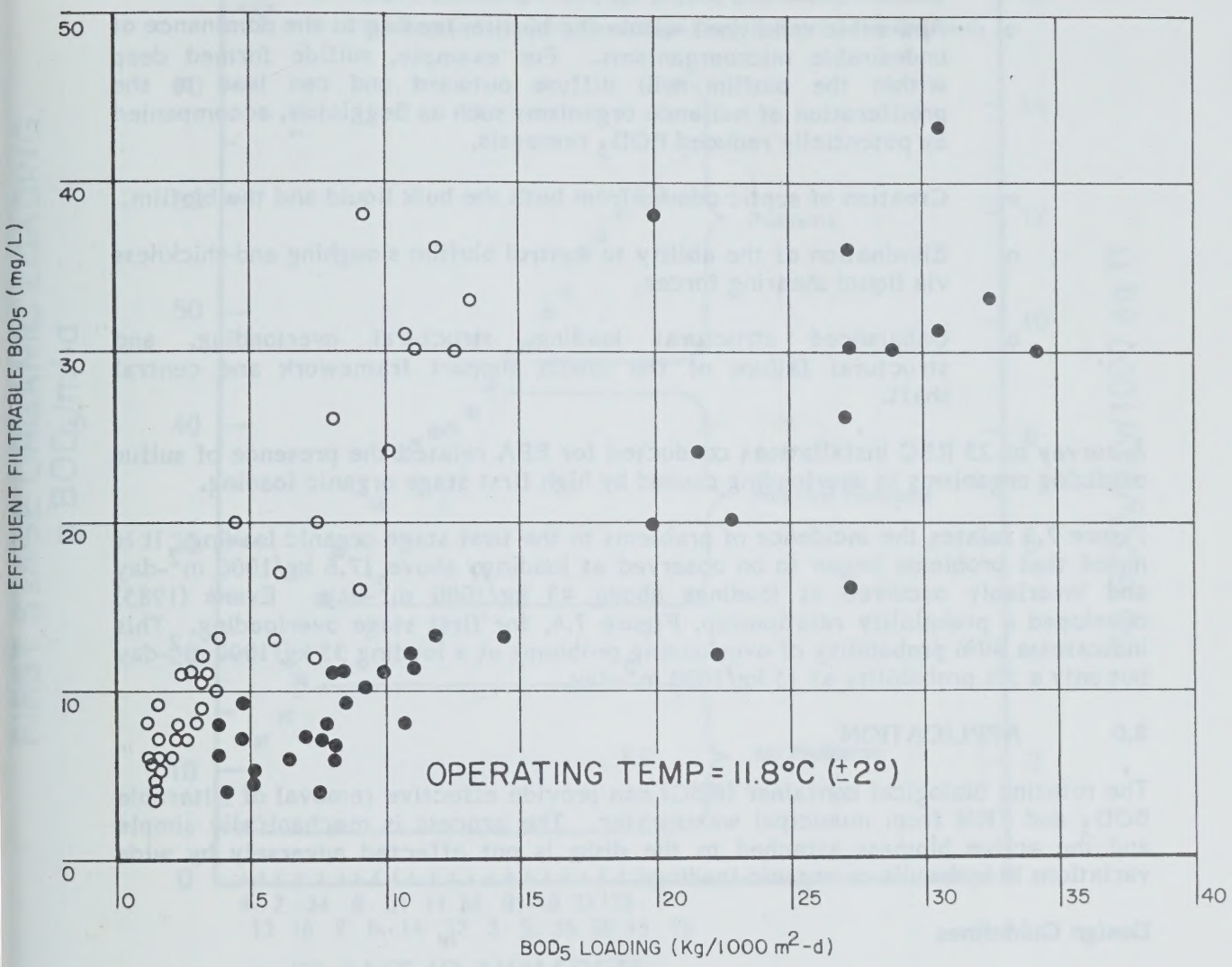


FIGURE 7.2 BOD_5 Loading vs. Filtrable Effluent BOD_5 (2.0 m RBC) (After Murphy and Wilson, 1980)

WITHOUT PRIMARY CLARIFICATION

● TOTAL BOD_5 LOAD

○ FILTRABLE BOD_5 LOAD

- o Anaerobic conditions within the biofilm leading to the dominance of undesirable microorganisms. For example, sulfide formed deep within the biofilm will diffuse outward and can lead to the proliferation of nuisance organisms such as Beggiatoa, accompanied by potentially reduced BOD_5 removals.

- o Creation of septic odours from both the bulk liquid and the biofilm.

- o Elimination of the ability to control biofilm sloughing and thickness via liquid shearing forces.

- o Unbalanced structural loading, structural overloading, and structural failure of the media support framework and central shaft.

A survey of 23 RBC installations conducted for EPA related the presence of sulfur oxidizing organisms to overloading caused by high first stage organic loading.

Figure 7.3 relates the incidence of problems to the first stage organic loading. It is noted that problems began to be observed at loadings above 17.6 kg/1000 m^2-day and invariably occurred at loadings above 43 kg/1000 m^2-day. Evans (1985) developed a probability relationship, Figure 7.4, for first stage overloading. This indicates a 50% probability of overloading problems at a loading 35 kg/1000 m^2-day but only a 5% probability at 15 kg/1000 m^2-day.

8.0 APPLICATION

The rotating biological container (RBC) can provide effective removal of filterable BOD_5 and TKN from municipal wastewater. The process is mechanically simple and the active biomass attached to the disks is not affected adversely by wide variations in hydraulic or organic loadings.

Design Guidelines

"Mass Loading" has been found to be the most suitable design parameter to adopt as a basis for sizing RBC systems. For removal of BOD_5 from wastewater the mass loading is expressed as the mass of filterable BOD_5 applied per unit disk area and time (e.g., kg BOD_5/1000 m^2-d). For combined BOD_5 removal plus nitrification the loading is expressed as the mass of filterable TKN per unit disk area and time (e.g. kg TKN_F/1000 m^2-d). Recommended design loadings appear in Table 8.1. These loadings have been selected based on a full scale, short hydraulic detention, RBC system experiencing a peak to mean to minimum diuranl flow of 2:1:0.5. BOD_5 loadings which are 20 percent higher and TKN loadings which are 30 percent higher than those indicated in Table 8.1 could be applied if full equalization of the wastewater is provided. The loadings in Table 8.1 apply only to primary treated municipal wastewater. An allowance has been made for reduced treatment efficiency at lower temperatures. BOD_5 and suspended solids concentrations listed under "Design Objectives" represent the average, not the maximum, anticipated effluent concentrations at the indicated loadings.

Table 8.1 indicates that a loading of 6.0 - 6.5 kg total BOD_5/1000 m^2-d (1.2 - 1.3 lbs BOD_5/1000 ft^2-d) would be recommended to meet an average treated effluent objective of 15 mg/l each for BOD_5 and suspended solids assuming a minimum wastewater temperature of 8 - 10°C. This loading is 20 - 30 percent higher than

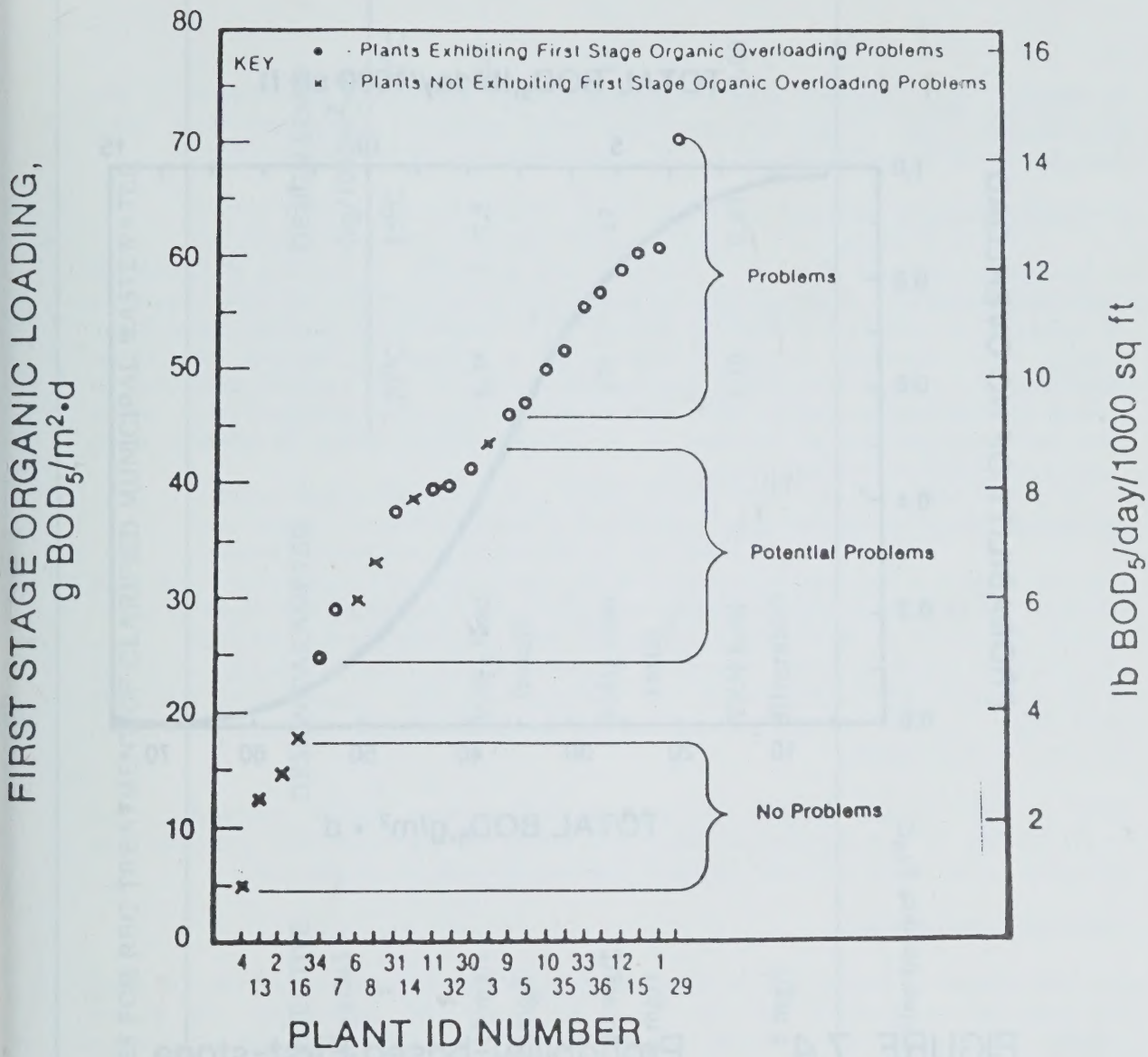


FIGURE 7.3 Array of First-stage Organic Loading Data (After Evans, 1985)

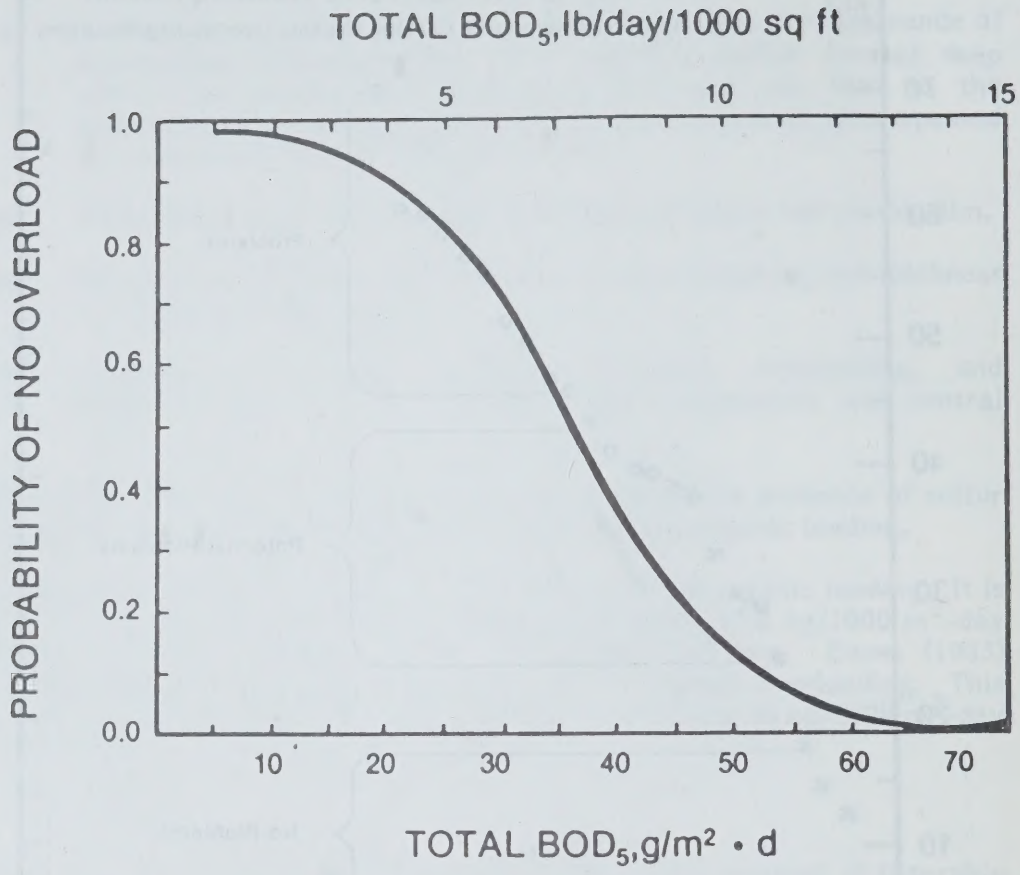


FIGURE 7.4 Probability-based First-stage Organic Loading Values for Design (After Evans, 1985)

TABLE 8.1: DESIGN GUIDELINES FOR RBC TREATMENT OF CLARIFIED MUNICIPAL WASTEWATER

RBC	DESIGN OBJECTIVE (average value)	DESIGN PARAMETER	DESIGN LOAD ($kg/1000m^2$-d)			
			20°C	15°C	10°C	5°C
BOD_5 Removal	BOD_5 15 mg/l SS 15 mg/l	BOD_5 load (total)	8.5*	8.5	6.5	5.0
BOD_5 Removal	BOD_5 30 mg/l SS 30 mg/l	BOD_5 load (total	17*	17	13	10
BOD_5 Removal Plus Nitrification	TKN 5 mg/l	TKN load (filterable)	1.10	0.65	0.40	0.25

* Assumes no further increase in removal rates beyond 15°C

Note: 1 lb/1000 ft^2-d = 4.89 kg/1000 m^2-d.

the design guideline for RBC's established by the Ontario Ministry of the Environment but is substantially less than loadings calculated from RBC manufacturers' design manuals.

Performance under Transient Loading

A hydraulic residence time of 3 hours or less depending on loading is typical of many full scale RBC systems. Under these conditions the RBC has little capacity to equalize or absorb variable or shock loads. Significant changes in effluent quality were observed within 30 minutes of a change in influent conditions. Significant variations in effluent quality will occur in RBC systems subjected to normal diurnal variations in flow and in BOD_5 and TKN concentration unless additional equalization is provided or a suitable design factor is applied when calculating surface area requirements.

Suspended Solids Characteristics

Treated effluents containing filterable BOD_5 concentrations consistently less than 10 mg/l can be produced by RBC's. Under these conditions the final effluent quality measured as total suspended solids and total BOD_5 was limited by the efficiency of the final solids separation process. Some of the settleable suspended solids in treated RBC effluent had low settling velocities. Conservative overflow rates (e.g., 1.0 m/h) should be selected for the final clarifier to obtain maximum removal of these solids.

RBC systems should include a process for primary solids removal. The concentration of non-settleable suspended solids in clarified RBC effluent was consistently less when primary clarified wastewater was fed to the RBC.

Scale-Up Considerations

Possible effects of process scale-up can be based on the comparison of a 0.5 m RBC was operated in parallel with the 2.0 m diameter RBC by Murphy and Wilson (1980). Under identical operating conditions the 0.5 m RBC averaged 16 percent higher mass removals of COD per unit area and time than the 2.0 m unit. The difference in removals was statistically significant at a confidence level of 95%. Under combined BOD_5 removal plus nitrifying conditions mass removals of filterable TKN were similar for both units. This would indicate that the oxygen transfer capability of the 2.0 m unit may be limiting at high loadings.

Full sized RBC shafts are generally 3.5 m in diameter. When designing for BOD_5 removal it would be recommended that a scale-up factor of 25% be applied to 0.5 m RBC pilot data. A 10% factor is recommended for scale-up from 2.0 m RBC test data. No scale-up factor appears to be required for combined BOD_5 removal plus nitrification systems.

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Pilot Scale Comparison of Activated Sludge and RBC Units For Mixed Domestic and Food Processing Waste

By

Michael N. Lywood (1)
Richard V. Laughton (2)

INTRODUCTION

The Regional Municipality of Niagara (slide 1), in the process of upgrading the existing 12.8 MGD Niagara Falls primary treatment plant to a secondary treatment plant, undertook a pilot plant study to determine the best form of secondary biological treatment. The plant upgrading to secondary treatment was proposed by the Region and was not a requirement of the Ontario Ministry of the Environment. Although the full scale plant construction is now complete and a one year test program is underway to determine the precise operating requirements of the full scale plant, this paper deals only with the pilot scale evaluations which led to the selection and final design of the RBC (Rotating Biological Contactors) process.

The Niagara Falls W.P.C.P. has an abnormal waste in that it is on a combined sewer system which results in very high dilution, therefore very low organic loading during the Spring, while during the Fall the dilution is minimal and food processing plants are discharging to the plant resulting in very high organic loadings to the facility. During the food processing season there are several abrupt changes in the feed characteristics entering the plant as a result of the changes in wastes that are entering the plant from the several food processors, wineries and canners in the area.

The Regional Municipality of Niagara retained the services of Pollutech Limited to undertake a detailed laboratory and pilot scale test program to determine the precise requirements for secondary biological treatment at the Niagara Falls facility. During the test period a pilot scale RBC and a pilot scale Activated Sludge unit were operated side-by-side at the Niagara Falls plant. Concurrently 3 lab scale activated sludge units were run at different retention times to evaluate the variations in hydraulic retention time which would not have been possible with the pilot scale facility. Additional on-site testing was also carried out to determine capacity of the primary clarifiers and anaerobic digestion facility. Chemical testing of the feed to the plant was undertaken to determine whether chemical pre-treatment could be used to minimize the size and cost of a secondary treatment facility.

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EXISTING FULL-SCALE PLANT

The Regional Municipality of Niagara Falls Water Pollution Control Plant Water Pollution Control Plant is currently rated for an average flow of 12.8 MGD and a peak flow of 25.6 MDG. The plant consists of six primary clarifiers having a total area of 33,240 ft^2 and a total volume of 349,000 ft^3. The overflow rate for these primary clarifiers at the average flow is 650-750 gallons/day/ft^2. At the time of this study there were no secondary treatment facilities at this plant and the primary effluent was chlorinated and discharged via the Hydro canal to the Sir Adam Beck Generating Station and then to the Niagara River. Some 25,000 gal/day of primary sludge was handled by two primary anaerobic digesters and one secondary anaerobic digester which provided 30 days of primary sludge digestion at a loading rate of 0.08 lbs VSS/day/ft^3.

The Regional Municipality of Niagara undertook to upgrade the plant to provide 15 MGD treatment on average for a population of 107,050. Ultimately they wish to provide 22 MGD peak secondary treatment and 30 MGD peak primary treatment. Testing was to be conducted to determine which form of secondary treatment could best be utilized and what were the capacities of the existing primary clarifiers and sludge digestion facilities.

DETAILED TEST PROGRAM

The major design considerations (slide 2) that were specified in the very early stages of this program were as follows:

- There was a very heavy load of food processing waste entering this plant during the late summer and fall amounting to some 14,000 pounds/day of BOD_5;
- During the spring period the combined sewers in this area routinely discharged in excess of 30 MGD;
- What ever treatment was to be considered must provide adequate cold weather treatment anticipating 5°C to 8°C winter temperatures.

The above three points were of concern in that the two major contenders for the secondary treatment were an activated sludge system and rotating biological contactors. Since these two processes react differently to rapid swings in hydraulic or organic loadings, as well as cold weather, it was essential to find the limits of the treatment process prior to undertaking any of the detailed design.

In the late summer of 1983 pilot scale treatment units were moved into the Niagara Falls Water Pollution Control Plant provide simulation of activated sludge and RBC waste treatment. A pilot scale activated sludge system that had been used many times before by Pollutech was installted along with a series of bench-scale activated sludge units that could be utilized to model variations in hydraulic retention time at the same time the pilot scale testing was ongoing. Along with the activated sludge system a household type RBC was installed as available commercially from CMS Rotordisk in Mississauga (slide 3). This RBC unit was a four chamber 500 sq. ft. unit (slide 4).

Throughout the pilot scale testing of the biological processes full-scale testing of the primary clarifiers was ongoing. The full-scale primary clarifier testing involved splitting the existing plant in half by way of a diversion installed in the primary clarifier feed channel (slide 5). One-half of the primary clarifiers were operated under current practices of adding ferric chloride and the other half was operated without any chemical treatment. Additionally, variations in sludge blanket levels and hydraulic flows to the clarifiers were studied in detail. The results of the primary clarifier tests are not discussed in detail in this paper. Primary effluent from either the chemically treated raw sewage or the straight primary effluent could be pumped to the lab scale and pilot scale test units.

One of the major advantages of this particular test program was the ability to opreate the activated sludge and RBC units in a side-by-side real life comparison (slide 6). The bench-scale activated sludge units were utilized such that we could evaluate at least three variations of hydraulic retention time for the activated sludge unit. All three of the test apparatus were installed in the gallery of the Niagara Falls Water Pollution Control Plant and all were fed on a continuous loop of primary effluent from the full-scale primary clarifiers.

Throughout the test period composite samples were collected on the raw sewage, two primary effluents, RBC effluent, activated sludge effluent and the bench-scale test effluents several times per week. Each sample received detailed analysis in Pollutech's Oakville laboratory.

The major impact on the operation of this water pollution control plant was the industrial and domestic variation in hydraulic and organic loading (slide 7). For the existing total plant flow of approximately 12.6 MGD, approximately 12 MGD was domestic flow and 0.6 MGD was industrial flow. However, at an industrial waste concentration of approximately 2200 mg/L as compared to 80 mg/L for the domestic sewage the organic load from the industrial operations amounted to some 13,000 - 14,000 lbs/day of BOD_5 as compared to only 9,600 lbs/day of BOD_5 for the domestic operations. The plant influent had a resultant BOD concentration of approximately 180 - 190 mg/L representing in excess of 23,000 lbs/day of BOD_5. An outline of the industrial and domestic input from the existing 12 MGD to the anticipated 20 MGD flows is provided in the attached Table 1.

SECONDARY INFLUENT CHARACTERISTICS

The primary treatment facilities were evaluated either with or without ferric chloride addition depending on the season (slide 8). Currently, ferric chloride is added at the front of the primary plant as part of the phosphorus removal program. The effect of the ferric chloride addition was quite substantial (slide 9) and can be summarized as follows:

- Raw sewage BOD ranged between 75 mg/L and 390 mg/L with an average of 173 mg/L and was equal to or less than 250 mg/L 90 percent of the time;

- Raw sewage Suspended Solids ranged between 122 mg/L and 1060 mg/L with an average of 261 mg/L and was equal to or less than 404 mg/L 90 percent of the time;

- Primary Effluent BOD_5 without chemical treatment ranged between 58 mg/L and 271 mg/L with an average of 132 mg/L and was equal to or less than 185 mg/L 90 percent of time;

- Primary Effluent Suspended Solids without chemical treatment ranged between 40 mg/L and 182 mg/L with an average of 92 mg/L and was equal to or less than 120 mg/L 90 percent of the time;

- Primary effluent BOD_5 after ferric chloride treatment ranged between 35 mg/L and 207 mg/L with an average of 113 mg/L and was equal to or less than 130 mg/L 90 percent of the time;

- Primary effluent Suspended Solids after ferric chloride treatment ranged between 14 mg/L and 372 mg/L and was equal to or less than 156 mg/L 90 percent of the time;

- The primary clarifier effluent Suspended Solids decreased from a high of 372 mg/L down to 73 mg/L through implementation of a solids removal stategy, while under similar hydraulic loading conditions and with ferric chloride treatment;

Summary data on the flow, BOD_5, and suspended solids is provided for 1981 through 1983 in the attached Table 2. Specific information during the three month test period is provided in Table 3 showing both results with and without ferric chloride treatment.

CONCEPTUAL PLANT DESIGN

The process design of the facilities included allowances for the following items to optimize tank requirements and ensure effluent criteria could be met:

- Ferric chloride could be used during the processing season to shave the peaks off the organic loading;

- The fraction of industrial waste in the plant influent would not likely increase significantly.

The field test results for the primary clarifiers show an average 20 percent BOD_5 removal without ferric chloride addition, and 40 percent removal with ferric chloride addition. This did not change significantly for flows up to 20 MGD.

For design of the activated sludge process it was deemed essential to design a system capable of handling the high organic loads, otherwise the system would fail. Using the 90 percent less than or equal to figure, the primary effluent BOD_5 data would be as follows:

Plant Flow (MGD)	Primary Effluent BOD_5 (mg/L)	Primary Effluent BOD_5 (lb/Day)
12.6	204	25,704
15	187	28,050
20	174	34,800

Test results showed the raw sewage to have a BOD_5 of 250 mg/L or less on 90 percent of occurrences. With 20 percent BOD_5 removal across the primaries this would give a primary effluent BOD_5 of 204 mg/L. In general, the 90 percent BOD_5 value is 40 percent greater than the expected average primary effluent BOD_5.

As stated previously, the design criteria had to be based on loadings experienced during the heavy seasonal processing period, not on the annual average. It should also be noted that during the seasonal processing period the organic matter (BOD_5) is not typical of domestic sewage. Food waste organics are known to biodegrade more slowly, and with different microbial populations, than human organic wastes.

With the seasonal use of ferric chloride to control the BOD_5 the expected conditions going to secondary treatment would be:

Plant Flow (MGD)	Primary Effluent BOD_5 (mg/L)	Primary Effluent BOD_5 (lb/Day)
12.6	126	15,876
15	115	17,250
20	107	21,400

Test results showed the primary effluent BOD_5 with ferric treatment was less than 130 mg/L on 90 percent of occurrences. On the average, the 90 percent value is 15 percent higher than the average primary effluent BOD_5.

Unlike the plant loadings for the activated sludge process, the RBC system could be based on the average primary effluent quality, rather than the 90 percent less than value. This is due to the demonstrated ability of the RBC units to accept short-term peak organic loadings without any deterioration in the effluent quality.

The design values are thus:

Condition	Plant Flow MGD	BOD_5 (mg/L)	BOD_5 (lb/day)
Without Ferric	12.6	147	18,520
	15	134	20,100
	20	124	24,800
With Ferric	12.6	110	13,860
	15	100	15,000
	20	93	18,600

The loading data for the RBC system is based upon field test data showing 20 percent BOD_5 removal across the primaries without ferric chloride addition, and 40 percent BOD_5 removal with ferric chloride addition. The average raw sewage BOD_5 data is that used for projections up to 20 MGD, as previously outlined for the activated sludge process.

PILOT SCALE TESTING

The detailed pilot scale test included the operation of three bench-scale aeration systems in which we evaluated hydraulic retention times of 5 hours, 7 hours, and 10 hours (slide 10).

Operation of the bench-scale aeration systems clearly indicated there was a significant variation in the operation prior to and during the food processing season and that hydraulic retention times suitable of providing a good effluent quality and a well settling sludge under routine operating conditions could not provide similar operations during the food processing season (slide 11). During the food processing season the hydraulic retention time of at least 24 hours would be required prior to having any hope of maintaining the operation of an activated sludge system. This long hydraulic retention time would be totally unsuitable during the remainder of the year during which time 6 hours aeration time would be optimum. On an operating sense this provides a major problem in co-ordinating how many aeration cells would be in use. The variations within the food processing season was also significant, indicating very close operating attention would have to be provided from late summer through to early winter to maintain adequate operation of an activated sludge system.

In conjunction with the bench-scale treatability test for the activated sludge system a pilot scale activated sludge system was maintained in operation throughout the food processing season at a hydraulic retention time of approximately 7 hours with a solids retention time of approximately 10 days (slide 12).

Detailed 8 foot tube settling tests (slide 13) were conducted on the sludge from the pilot scale activated sludge system and after over 60 minutes of settling less than 6 inches of clear effluent was present in the 8 foot settling column (slide 14).

The pilot scale rotating biological contactor was set up such that primary effluent was fed into the first stage and effluent could be taken out at either the third or fourth stage of the RBC test unit. The household unit did not have a secondary clarifier that would be typical of a full-scale application and therefore, we modified a 45 gallon steel drum to simulate a typical clarifier operating in conjunction with the RBC unit (slide 15).

The RBC pilot system operated successfully throughout the summer season prior to the food processing season and also throughout the summer/fall food processing season. Little attention was required to the RBC unit as compared to the pilot scale activated sludge system. Our major operating control parameters were found to be control of the dissolved oxygen levels in the four RBC chambers and monitoring of the soluble BOD_5 removal in each of the four chambers (slide 16).

Operation of the pilot scale RBC system showed that an effluent quality ranging from 7 mg/L to 20 mg/L could be attained and organic loading rates ranging from 1 lb/1000 ft^2/day to 4 lb/1000 ft^2/day as illustrated in attached Table 4 (slide 17). If organic loading rates were allowed to increase much beyond 4 lbs/1000 ft^2/day a thick white growth of Beggiatoa (slide 18) became predominant on the front disk of the pilot unit.

The RBC process design uses an organic loading of 2.2 lb BOD_5/1000 ft^2/day. The value is a conservative number based on an acceptable pilot plant loading of 3.2 lb BOD/1000 ft^2/day, which was adjusted for increased disc diameter, operation under variable flow conditions, and cold winter conditions. The various disc surface area requirements would be:

Condition	Plant Flow (MGD)	Disc Area (ft^2)
Without Ferric	12.6	8,420,000
	15	9,140,000
	20	11,270,000
With Ferric	12.6	6,300,000
	15	6,820,000
	20	8,435,000

During the field test program we observed that the effluent quality leaving the third compartment of the RBC was as acceptable as effluent leaving the fourth compartment. Utilizing this data suggests that under maximum loading conditions the RBC could accept a pilot plant loading of 4.1 lb BOD_5/1000 ft^2, equivalent to 2.8 lb BOD_5/1000 ft^2 at full-scale.

In order to check the limiting operating period of the RBC we calculated the surface area requirements under the peak food processing season, and also during the average cold weather conditions. A comparison of these two sets of data shows the following:

Conditions	Plant Flow (MGD)	Total Disc Area (million ft^2)	
		Winter (8°C)	Summer (20°C)
Without Ferric	12.6	5.8	8.4
	15	6.8	9.2
	20	9.1	11.3
With Ferric	12.6	4.5	6.3
	15	5.4	6.8
	20	7.2	8.5

Please note the organic loading is not constant between the winter and summer. This shows the summer conditions to be limiting due to organic loading. If however the winter temperature dropped to 5°C or lower, then the winter design criteria becomes critical.

The RBC process does not result in the serious sludge settling problems from a filamentous culture, as was found in the activated sludge process. The peculiarity of the RBC clarifier is that it must be sized on the basis of settling discrete particles rather than flocculant particles. The field test data showed that the loss of small particles (less than 37 micron) could seriously affect the effluent quality. Tests were conducted to evaluate the particle size of the solids escaping from the clarifier, to determine best how to maintain the solids in the system at high flows (slide 19).

Tests conducted during the latter parts of the study indicated that clarifier operation could be improved by taking the feed to the clarifier from the third compartment, adding a small amount of cationic polymer, and optimizing sludge blanket control in the clarifier (slides 20, 21).

There has been little work conducted on the sizing of RBC clarifiers. Most references refer to sizing on the basis of conventional parameters. Most US plants however do not have 15/15 effluent regulations. To be successful we feel the clarifiers must be sized more conservatively.

During the pilot test program the RBC system was evaluated using a clarifier having an effective surface area of 2.6 ft^2. Hydraulic loading to this clarifier ranged from 222 gpd/ft^2 to 1330 gpd/ft^2. The test results show that as long as the sludge blanket is controlled in the bottom of the clarifier the unit is resiliant to hydraulic peaks. Since the particles settle under discrete rather than flocculant conditions, the RBC clarifier more closely resembles a conventional primary clarifier than a conventional secondary clarifier.

Information supplied by the RBC manufacturers does suggest that clarifier operation improves with an increase in total size, rather than deteriorating. We have also observed a similar relationship on previous pilot work we have conducted for discrete settling particles.

Utilizing all of the information available, the requirements for a full-scale RBC clarifier system based on an average overflow rate of 300 gpd/ft^2 or 600 gpd/ft^2 peak flow would be:

Plant Flow (MGD)	Peak Flow (MGD)	Total Clarifier Area (ft^2)
12.6	25	42,000
15	30	50,000
20	40	66,700

Minimum liquid depth in the clarifier should be 10 feet, and current technology regarding weir placement for control of density currents should be practiced.

FINAL PROCESS DESIGN

The data obtained from the pilot plant program could be extrapolated to produce design criteria for each process component required to meet the effluent objective. However, there are several sets of design possibilities, because of alternate strategies available for operating criteria for the plant. For instance, there are:

- two flow capacities limits (i.e. 15 MGD and 20 MGD);
- two alternatives for primary clarification (with chemical treatment or without);
- two biological process alternatives (activated sludge and rotating biological contactor);
- two strategies for handling maximum wet weather flows (all of it through the plant, or partial bypassing of secondary system).

Thus it is readily apparent that for each of these factors with two alternatives there are sixteen combinations of design criteria. For the preliminary process design the Region and Pollutech developed a number of assumptions in order to develop design criteria for the two most probable options for the plant:

- 15 MGD average daily flow;
- comparable activated sludge and RBC's for capital and operating costs;
- operation of the primary clarifiers with chemical treatment during the season of strong wastes, even though this may not be the principal strategy for phosphorus removal it is needed to shave the peaks off the BOD_5 and Suspended Solids;
- a case can be made to the Ministry for bypassing of wet weather peaks exceeding 22 MGD around the biological section.

Under this set of conditions the following are the design criteria which would apply based on the pilot plant results:

- primary clarifiers will not require increased tankage;
- either - activated sludge with an active capacity of 7.7 million gallons and an oxygen capability of 1085 pounds per hour and a final clarifier of 58,800 sq. ft.
  or - an RBC of 6,820,000 sq ft (including the manufacturers scale up criteria) plus a final clarifier of 50,000 sq. ft.

The rationale for selecting the disk area requirements was based on plant flows of 12, 15 and 20 MGD, having treatment with and without ferric chloride, and evaluating both summer and winter conditions. This information is as shown on Table 5 (slide 22).

The RBC design values which were acceptable at the pilot scale stage as well as practical at the full scale stage are shown in the attached Table 6 (slide 23). This information indicates that an acceptable pilot scale loading of 4.1 lbs. BOD_5 per 1000 $ft.^2$ could be reduced to as low as 2.2 lbs. BOD_5 per 1000 ft^2 for practical full-scale operation. A pilot plant operation at 3,600,000 ft^2 could require 6,800,000 ft^2 at full scale given scale up factors for disc size and temperature.

The Regional Muncipality of Niagara undertook to proceed with the first stage of the RBC system by putting in approximately half the necessary disk area or 3.5 million square feet. This size of plant could be funded within the $12,000,000 budget. In the event that the full-scale system operated in a similar manner to the pilot scale system, that is without any regard for scale up factors, then it would be possible that this amount of disk surface area could suffice during many periods of the year. The Ontario Ministry of the Environment issued a Certificate of Approval on the condition that by 1990 the industrial load was reduced sufficiently to match the plant capacity, or the Region would have to expand the facilities. The Region is now implementing the cooperative approach with local industries.

The rationale for ultimate treatment at the plant was therefore based on the installation of 3,500,000 ft^2 of disk surface area with a certain amount of the flow being treated by the complete primary/secondary system in parts of the year and at other times of the year having flow bypass the secondary treatment operation. The details of the rationale area provided in the attached Table 7 (slide 24) which shows the percent of the flow that would require primary and secondary treatment at various times in the year and the projected effluent quality of the mixed streams. Since it was not mandatory for the Niagara Falls Stanford Water Pollution Control Plant to go to secondary treatment it was not essential to meet the standard 15:15 effluent quality. Ultimately what was decided was to attain 15:15 effluent quality during warm weather period without industrial inputs from the food processing plants, but allow effluent quality as high as 63 BOD_5 and 47 solids during the food processing season. If the operation of the RBC's was better than anticipated an effluent quality of 40:30 could be achieved during this period.

The full-scale facilities have now been constructed and a 18 month start-up and test program is being carried out by Pollutech Limited in close cooperation with Regional operation's staff. The start-up of the system did not begin prior to the heavy food processing season in the fall of 1985, therefore, it will be some months yet before the true capabilities of the RBC units can be identified.

SUMMARY

The results of the laboratory and pilot scale test program clearly illustrated that the activated sludge system could not adequately respond to the dramatic changes in feed quality experienced during the test period. During this same period the RBC system responded with only minimal impact from the abrupt changes in feed concentration. Final selection of the RBC system as the most applicable secondary process was <u>not</u> a matter of comparing effluent BOD_5 and Suspended Solids values, but rather an observation of the inability of the activated sludge system to survive the waste loadings.

With substantial grants from both the Federal and Provincial Governments, the Region has been able to implement a program for secondary effluent treatment for the City of Niagara Falls, despite harsh conditions arising from heavy hydraulic and organic loadings. The process will utilize combined physical/chemical/biological treatment to provide the best blend of treated effluent during various seasonal periods.

TABLE 1 - SOURCES OF PLANT HYDRAULIC AND ORGANIC LOAD

Conditions	Industrial	Domestic	Total
Existing 12.6 MGD	0.62 MGD 2188 mg/L 13,566 lb/day	12.0 MGD 80 mg/L 9,600 lb/day	12.62 MGD 183 mg/L 23,166 lb/day
Increase to 15 MGD	0.62 MGD 2188 mg/L 13,566 lb/day	14.4 MGD 80 mg/L 11,520 lb/day	15 MGD 167 mg/L 25,086 lb/day
Increase to 20 MGD	0.62 MGD 2188 mg/L 13,566 lb/day	14.4 MG 80 mg/L 11,520 lb/day + 5.0 MG 120 mg/L 6,000 lb/day	20 MGD 155 mg/L 31,086 lb/day

TABLE 2 - PLANT OPERATIONS DATA

	1981	1982	1983 (Jan.-Oct.)
Flow (MGD)			
Average	9.8	11.4	12.0
Maximum Day	22.5	24.6	23.5
Average BOD_5(mg/L)			
Raw Sewage	102	101	97
Primary Effluent	27	42	33
Average Suspended Solids (mg/L)			
Raw Sewage	164	179	173
Primary Effluent	26	26	21

TABLE 3 - TEST PERIOD DATA

	Aug.	Sept.	Oct.	Avg.
Flow (MGD)				
Average	13.3	12.4	12.8	12.8
Maximum Day	19.3	16.8	21.3	21.3
Average BOD_5 (mg/L)				
Raw Sewage	141	192	186	173
Primary no Ferric	93	146	157	132
Primary with Ferric	94	130	114	113
Average Suspended Solids (mg/L)				
Raw Sewage	183	343	258	261
Primary no Ferric	99	88	90	92
Primary with Ferric	90	120	34	81

TABLE 4 - RBC LOADING RATES

Pilot Disc Organic Loading (lb/1000 ft^2/day)	Pilot Disc Effluent BOD_5 (mg/L)
1.0	7
2.0	11
3.0	14
4.0	20

TABLE 5 - RBC DISC AREA REQUIREMENTS

Plant Flow (MGD)	WITH FERRIC			WITHOUT FERRIC		
	BOD_5 lb/day	Disc Area (10^6 ft^2)		BOD_5 lb/day	Disc Area (10^6 ft^2)	
		Summer	Winter		Summer	Winter
12	18,500	8.4	5.8	13,900	6.3	4.5
15	20,100	9.2	6.8	15,000	6.8	5.4
20	24,800	11.3	9.1	18,000	8.5	7.2

TABLE 6 - RBC DESIGN VALUES (15000 lb/day BOD_5)

	lb/BOD_5/1000 ft^2	DISC AREA 10^6 ft^2
Acceptable at Pilot Stage	4.1	3.6
Practical Pilot Date	3.2	4.7
Acceptable Full Scale	2.8	5.4
Practical Full Scale	2.2	6.8

TABLE 7 - RATIONALE FOR PARTIAL TREATMENT

Condition	Percent of Flow		Effluent (mg/L)	
	Primary Only	RBC's Also	BOD_5	Solids
2.2 lb/1000 ft^2 Warm Spring, No Industrial (April-June)	0	100	15	15
1.75 lb/1000 ft^2 Cold weather, No Industrial (Dec.-Feb.)	20	80	35	28
2.2 lb/1000 ft^2 Warm Fall, With Industry (Aug.-Oct.)	49	51	63	47
3.2 lb/1000 ft^2 Processing Season Possible (Aug.-Oct.)	26	74	40	32

ABSTRACT

NEW DEVELOPMENTS IN BIOLOGICAL FILTRATION

By

Denny S. Parker
Brown and Caldwell

The trickling filter, or biological filter, was a mainstay of waste treatment until the seventies when it lost favor to the activated sludge process. The recent development of the trickling filter/solids contact process has restored the biological filter's status in the eighties as an effluent can be produced as good as the activated sludge at lower cost. Today there are at least 30 TF/SC projects in the United States that have progressed at least to the design stage; eight projects are in operation. The development of the TF/SC process by Brown and Caldwell will be reviewed starting from its initial application at Corvallis, Oregon, for tertiary quality effluents ("10-10"). The process has since been tailored to meet a wide range in effluent qualities, including both typical secondary levels ("30-30") as well as to accomplish nitrification.

SELECTED REFERENCE LIST

for

NEW DEVELOPMENTS IN BIOLOGICAL FILTRATION

By

Denny S. Parker
Brown and Caldwell

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